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EFFECTIVE FILTRATION METHODS FOR SMALL WATER SUPPLIES

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16. ABSTRACT

A 2-year study was conducted of various simple water filtration systems potentially appropriate for high-quality surface waters serving small systems. A slow sand filter without coagulant and a direct, rapid filter with coagulant were operated in parallel. Direct filtration with and without flocculation were compared in parallel in one phase of the study; declining- and constant-rate filtration were compared in parallel in another phase. The study was designed to emphasize simple treatment systems for small supplies where operational skill and attention may be lacking. The systems were compared while monitoring turbidity, particle count, and coliform bacteria in the influent and filtered water.

Slow sand filtration was the most effective for particulate removal, but filter runs were as short as 9 days during algal blooms. If the raw water is consistently high in quality and land is available, the slow sand filter would be the simple system of choice. All three direct filtration systems studied were capable of meeting the 1-nephelometric-turbidity-unit (NTU) maximum contaminant level (MCL), except during the first hour of the filter cycle. Flocculation was beneficial to the filtrate quality and head loss in direct filtration, but it was detrimental to the terminal breakthrough. Declining-rate filtration did not improve the filtrate compared with constant-rate filtration.

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FOREWORD

The U.S. Environmental Protection Agency was created because of increasing public and government concern about the dangers of pollution to the welfare of the American people. Noxious air, foul water, and spoiled land are tragic testimonies to the deterioration of our natural environment. The complexity of that environment and the interplay of its components require a concentrated and integrated attack on the problem.

Research and development is that necessary first step in problem solution; it involves defining the problem, measuring its impact, and searching for solutions. The Municipal Environmental Research Laboratory develops new and improved technology and systems to prevent, treat, and manage wastewater and solid and hazardous waste pollutant discharges from municipal and community sources, to preserve and treat public drinking water supplies, and to minimize the adverse economic, social, health, and aesthetic effects of pollution. This publication is one of the products of that research and provides a most vital communications link between the researcher and the user community.

This study was designed to evaluate the efficiency of various simple filtration systems that could be applied to high quality surface waters that might contain the pathogenic cysts of *Giardia lamblia*. Slow sand filters and rapid direct filtration alternatives were compared. The results have important implications for designers and operators of small water supply systems.

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Director
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Research Laboratory

ABSTRACT

A 2-year study was conducted of various simple water filtration systems potentially appropriate for high-quality surface waters serving small systems. A slow sand filter without coagulant and a direct, rapid filter with coagulant were operated in parallel. Direct filtration with and without flocculation were compared in parallel in one phase of the study; declining- and constant-rate filtration were compared in parallel in another phase. The study was designed to emphasize simple treatment systems for small supplies where operational skill and attention may be lacking. The systems were compared while monitoring turbidity, particle count, and coliform bacteria in the influent and filtered water.

Slow sand filtration was the most effective for particulate removal, but filter runs were as short as 9 days during algal blooms. If the raw water is consistently high in quality and land is available, the slow sand filter would be the simple system of choice. All three direct filtration systems studied were capable of meeting the 1-nephelometric-turbidity-unit (NTU) maximum contaminant level (MCL), except during the first hour of the filter cycle. Flocculation was beneficial to the filtrate quality and head loss in direct filtration, but it was detrimental to the terminal breakthrough. Declining-rate filtration did not improve the filtrate compared with constant-rate filtration.

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ABBREVIATIONS AND SYMBOLS

APHA	American Public Health Association
ASL	Analytical Service Laboratory
asu	areal standard unit (400 square micrometers)
BW	backwash
° C	degree Celsius
cm	centimeter
CRF	constant-rate filter or filtration
DO	dissolved oxygen
DRF	declining-rate filter or filtration
es	effective size
ft	feet
FTU	formazin turbidity unit
G	root mean square velocity gradient
g	gram
gpm	gallons per minute
gpm/ft^2	gallons per minute per square foot
h	hour
I.D.	inside diameter
in.	inch
L	liter
M	molar concentration

MCL	maximum contaminant level
m	meter
m/h	meter per hour
mm	millimeter
μm	micrometer
mg	milligram
mg/L	milligram per liter
mgad	million gallons per acre per day
mgd	million gallons per day
min	minute
NTU	nephelometric turbidity unit
pH	negative log of hydrogen ion concentration
rpm	revolutions per minute
s	second
TDS	total dissolved solids
t	time
tu	turbidity unit
uc	uniformity coefficient
UFRV	unit filter run volume
vs	versus
$\bar{x}$	average value
σ	standard deviation

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SECTION 1

INTRODUCTION

Some communities served by protected upland surface water supplies presently provide no drinking water treatment except for disinfection. This practice is especially common for small communities. Such supplies may exceed the EPA maximum contaminant level (MCL) for turbidity (1 nephelometric turbidity unit, or NTU) in public water supplies during some seasons of the year. Furthermore, a number of waterborne disease outbreaks have resulted in such communities.

Such communities are faced with the need to construct and operate some form of water treatment that will consistently produce water that will protect the public health and meet the drinking water standards. For the small community, simplified treatment systems are needed that will require a minimum level of operator skill for effective operation yet provide a system that will ensure acceptable levels of treated water quality.

The research reported here was directed to this problem and recognized the following developments and concerns:

1. The concern over potential Giardia transmission by public water supplies.
2. The growing use of direct filtration for high-quality raw waters. Direct filtration either eliminates flocculation or reduces the flocculation detention time provided and eliminates sedimentation.
3. The recognition that the old slow sand filtration concept still has potential applications for high-quality raw waters and has the advantage of eliminating chemical pretreatment. The slow sand filter is a lower level of water treatment technology and thus offers the potential advantages of requiring less operational attention and less-skilled operators. Thus the application to small systems with adequate available land may be attractive.

This report presents the results of a 2-year pilot study of simplified filtration techniques applicable to small public water supplies treating high-quality surface waters. The results are subdivided into four main parts:

1. Results for a slow sand filter operated for 15 months without chemical pretreatment of any sort.

2. Results for a rapid, dual-media, constant-rate filter operated in the direct, in-line filtration mode using alum or cationic polymer as a sole coagulant.

3. An evaluation of the impact of flocculation on direct filtration obtained by a 3-month parallel operation of two constant-rate filters, one operated with flocculation before filtration, one operated without flocculation, and both with the same chemical pretreatment.

4. A parallel comparison of constant-rate filtration with declining-rate filtration while both systems were operated in the direct, in-line filtration mode.

In all portions of the study, raw and filtered water were monitored for turbidity, particle count, coliform bacteria, and head loss development. The raw water was a high-quality surface water in a gravel pit in Iowa. The study covered a full range of seasonal extremes, with water temperatures from 2° C in the winter under ice cover to 25° C in the summer. The summer season included several intense algal blooms.

SECTION 2

OBJECTIVES

The original specific objectives of the project were:

1. To compare the performance of slow sand filters with conventional rapid filters during direct filtration of surface water for the removal of particulates as measured by particle counting, turbidity, coliforms, and standard plate count. The slow sand filters would be operated without chemical pretreatment, and the rapid filters would receive water pretreated to destabilize the suspended solids.

2. To compare the performance of declining-rate and constant-rate filters when both are operated in the rapid, direct filtration mode using the same parameters of filter performance.

3. To focus attention on those portions of a filter run where filter performance would be subject to less than normal performance:

    - During the period immediately after backwashing the filter, when the filtered water quality first degrades and then improves to the level normally observed during the majority of the filter run, and

    - During the breakthrough period when the filtered water quality gradually deteriorates to unacceptable quality.

4. To use the results of the research to provide recommendations for design and operation of filters for small water supplies, not only to meet the turbidity standards of the Safe Drinking Water Act, but also to achieve the best possible filtrate.

SECTION 3

CONCLUSIONS

The following conclusions can be drawn from the various phases of the study.

SLOW SAND FILTRATION CONCLUSIONS

1. The filtrate quality was somewhat inferior for one to two days at the beginning of each filter run when compared with the quality for the remainder of the filter run. A period of filtering-to-waste of this duration is appropriate where Giardia cysts are of concern.

2. A gradual improvement occurred in the filter performance by four parameters (turbidity, particle count, total coliform bacteria, and chlorophyll).

   After the first four filter runs spanning an 8-month period, the performance in each subsequent run (excluding the first two days of each run) was excellent, as follows: (a) average turbidity removal for each run was 97.8% or better (Table 14), (b) 7- to 12-μm particle removal for each run was 96.8% or better (Table 15), (c) 1- to 60-μm particle removal was 98.1% or better except in one run with 92.8% removal (Table 16), (d) total coliform bacteria removal was 99.4% or greater, reaching 100% in one filter run (Table 17), and (e) average chlorophyll-a removal was 95% or better, even after the second filter run (Table 18).

3. Filter run length was generally rather short--41 days or less in 9 out of 10 complete runs (Table 13), all of which were terminated by a steeply accelerating head loss curve. A long run of 123 days was achieved only under winter conditions, when algal populations were reduced. During serious algal blooms, runs were as short as 9 days. Increasing available head loss would not increase these run lengths appreciably because of the exponentially increasing head loss curves.

4. Turbidity alone was not an adequate predictor of the expected probable run length. Algal enumeration or a surrogate measure of algal population such as chlorophyll are essential parameters for judging the acceptability of a raw water for slow sand filtration.

5. No evidence showed that the filter was clogging to any substantial depth. The initial head loss at the beginning of each cycle reached a steady level after the first two runs and did not get progressively higher. This absence of depth clogging was also confirmed by scanning electron microscope examination of the sand at several depths at the end of the project.

DIRECT, IN-LINE FILTRATION CONCLUSIONS

The following conclusions are based on the direct, in-line, rapid filtration studies reported here, using alum or cationic polymer as a coagulant.

1. An initial period of poorer filtrate quality existed in all filter runs, as evidenced by turbidity, 7- to 12-μm particle count data, 1- to 60-μm particle count data, and total coliform data. Peak turbidity during this period often exceeded 1 NTU. The period of initial improvement lasted several hours in some cases, although the worst effects were over in 1 h. Thus where *Giardia* cysts are of concern, a filtering-to-waste period would be appropriate.

2. When serious algal blooms were not in progress, alum dosages between 5 and 10 mg/L (as $Al_2(SO_4)_3 \cdot 18\ H_2O$) or cationic polymer (Cat-Floc T) dosages between 0.09 and 1.49 mg/L could treat raw waters with average turbidities of 8 NTU and peak turbidities as high as 16 NTU and produce (a) acceptable filtrate with average turbidities well below 1 NTU before breakthrough, and (b) reasonable filter run length (Tables 20 and 21).

3. During a period of heavy blue-green algal population with chlorophyll-a level of 130 mg/m^3, and with average turbidity of 20 in the raw water, prechlorination was essential to the reasonable success of the direct, in-line filtration process. Alum dosages up to 20 mg/L were used with filter cycles as short as 12 hours at 7.3 m/h. Without prechlorination, filtrate quality of less than 1 NTU could not be assured. Even with prechlorination, the 1-NTU limit was sometimes exceeded.

4. With low algae (chlorophyll-a less than 5 mg/m^3), the mean solids load for the filter media of this study was 1.9 Kg suspended solids applied per square meter of filter area per meter of head loss increase ($Kg/m^2/m$) when using alum, and 2.5 $Kg/m^2/m$ when using cationic polymer. With moderate algae, the value dropped to 1.1 $Kg/m^2/m$ when using alum and 1.8 $Kg/m^2/m$ when using cationic polymer.

   Based on these values, the following limits of average raw water turbidity were calculated to achieve 24-hour cycles at 7.5 m/h filtration rate with 2 m of head loss increase available (above initial clean filter system head loss).

	Average Turbidity NTU	Average Suspended Solids mg/L
During low algae		
Using alum	12	21
Using cationic polymer	16	28
During moderate algae		
Using alum	7	12
Using cationic polymer	11	20

Higher values for short periods during the filter run can be tolerated providing the average is not violated.

5. The percent removal of 7- to 12-μm particles after the first hour of the cycle was above 85% in all cases (Table 22), exceeded 90% in 8 of 11 cases, and was greater than 95% in 6 of 11 cases.

6. The percent removal of total coliform bacteria after the first hour of the cycle was greater than 86% in all cases (Table 23), greater than 90% in 4 of 10 cases, and greater than 95% in 2 of 10 cases.

7. The percent removal of 7- to 12-μm particles generally exceeded the percent removal of total coliform bacteria. This result might be expected because of the greater size of the 7- to 12-μm particle compared with typical bacterial size.

8. The performance of direct, in-line filtration was not impaired by cold water as low as 2° C. In fact, when the best raw water was treated during the winter ice cover, excellent filtrate and long filter runs were obtained.

9. The cationic polymer produced substantially longer filter cycles than alum but a slightly inferior filtrate, as judged by all three parameters. Run-length data are summarized in Table 24. Run length and filtrate quality comparisons for various coagulants are clouded by the fact that the comparison runs were sequential rather than parallel.

10. Selecting the optimum coagulant dose for direct, in-line filtration was difficult because of the variability of raw water quality. Overdosing with alum caused excessive head loss and early breakthrough. Overdosing or underdosing with cationic polymer resulted in poorer filtrate quality throughout the run.

11. Selecting the optimum dosage of cationic polymer was more difficult than selecting the optimum dosage of alum because of the variable raw water quality. The proper dosage of alum was easier to select

because it was much less sensitive to raw water quality than the cationic polymer dosage. While operating the filters at a particular dosage of alum between 5 and 10 mg/L, raw water turbidity changes from 2 to 20 NTU had practically no impact on the filtrate quality.

12. Selecting the optimum dose of cationic polymer was assisted by briefly halting the polymer feed (about 10 to 20 min) and observing the turbidity response. If the earlier dosage was too high, the filtrate improved briefly (as the dosage residual in the filter diminished) and then deteriorated as the residual disappeared. If the earlier dosage was too low, the filtrate began to deteriorate immediately upon cessation of polymer feed.

DECLINING-RATE VERSUS CONSTANT-RATE FILTRATION (PHASE II)

The following conclusions relate to the comparison of declining-rate filtration with constant-rate filtration when both systems were operated in parallel using the direct, in-line filtration mode, and using alum or cationic polymer as a sole coagulant.

1. No water quality advantage occurred for the declining rate operation in turbidity, particle count, or total coliform removal compared with constant rate operation. This conclusion contrasts with an earlier study [26], in which a significant qualitative advantage for declining-rate operation was reported while filtering water from a lime-softening plant.

2. Rate of head loss increase was the same for the constant- and declining-rate operation at either 7.70 or 13.35 m/h (3.15-5.46 gpm/ft^2). The manner of media compaction after backwashing was very important in ensuring comparable head loss results.

3. The highest flow rate in the bank of declining-rate filters always occurred in the cleanest filter just after backwash.

4. The effluent turbidity, particle count, and total coliform counts were higher at the beginning of the run during the initial improvement period for both the declining- and constant-rate filters. No substantial decrease occurred in average effluent turbidity when a filter-to-waste period was used that consisted of wasting all effluent at the beginning of the run until the turbidity dropped to 0.5 NTU. The use of a filter-to-waste period, however, would eliminate the slug of turbidity, coliform bacteria, and cyst-sized particles that pass through the filter during the initial improvement period.

5. The particle counts and coliform counts followed the turbidity trends very well. The continuous turbidity graphs provided the best representation of the filtration run results.

6. The periodic backwashes of the declining-rate filters resulted in degradation of effluent water quality for a time during and after the backwash was complete. This result would also be expected in a bank of constant-rate filters but it was not observed in this research because only one was in operation.

THE IMPACT OF FLOCCULATION ON DIRECT FILTRATION (PHASE III)

The provision of a short period of flocculation (14 min) at a medium power intensity (G = 56 s^{-1} at 18° C) had the following impacts on the direct filtration performance. These observations are based on parallel operation of two constant-rate filters using either alum or cationic polymer as a sole coagulant. In some alum coagulation runs, acid was used to adjust the pH to about 6.8. Both filters received water that had received identical chemical dosages. One filter received flocculated water, and the other operated as an in-line filter without flocculation.

1. The filter receiving flocculated water had a shorter initial improvement period, as evidenced by lower average effluent turbidity and particle count data during the first hour of the run for the filter with flocculation (Tables 36, 40, and 41). This result was less clearly demonstrated with cationic polymer and with the 7- to 12-μm particle data.

2. The average quality of the filtrate during the remainder of the filter run, after the first hour and before terminal breakthrough, was superior for the filter with flocculation, as evidenced by turbidity and particle count data (Tables 36, 40, and 41). Again, this was less clearly demonstrated for all parameters when cationic polymer was used.

3. The provision of flocculation reduced the rate of head loss buildup when either alum or cationic polymer was the coagulant (Table 42). But in many alum runs, with or without pH adjustment, flocculation caused earlier breakthrough of turbidity (Tables 35 and 37).

4. When terminal breakthrough was a problem (as it was in many alum-coagulated filter runs), the lower head loss of the filter receiving flocculated water was of no benefit to the run length because the effective run length was controlled by breakthrough rather than by available head loss.

5. The other conclusions for direct, in-line, rapid filtration stated in Phases I and II were reinforced and supported by the work in Phase III.

GENERAL CONCLUSIONS RELATED TO SLOW AND RAPID DIRECT FILTRATION

The following general conclusions are drawn from the results and from the operational experience of the study:

1. The slow sand filter system studied in this research outperformed the direct, rapid filtration system operating with alum or cationic polymer as a primary coagulant. This conclusion was substantiated by turbidity, particle count, and total coliform bacterial data.

2. Where simple operation is important (as in small water supply systems), a slow sand filter system is superior to a direct, rapid filtration system, but the raw water must be of consistently high quality and low in algae to avoid excessively short cycles for the slow sand filter. Turbidity alone was not an adequate predictor of the probable run length. Algal enumeration or a surrogate measure of algal population, such as chlorophyll, are essential parameters for judging the acceptability of a raw water for slow sand filtration. Chlorophyll-a concentrations of less than 5 mg/m^3 along with turbidities of 5 NTU or less are suggested as upper limits for slow sand filter application.

3. For waters of somewhat poorer quality, rapid direct filtration can be used, but it requires substantially more operational skill and attention and poses a greater potential risk if improperly operated. Other alternatives such as diatomaceous earth filtration should also be considered.

4. The collection of raw water data on turbidity, suspended solids, and chlorophyll-a over a period of at least one year and including all seasonal extremes would be essential to make rational decisions among filtration alternatives.

5. Both slow sand filtration and rapid, direct filtration exhibited a period of poorer filtrate quality at the beginning of the filter runs. Thus, both systems require a filtering-to-waste period where Giardia cysts are of concern. Minimum wasting periods of 2 days for slow sand filtration and 1 h for rapid, direct filtration are suggested from the results of this study.

   Because of the need for a filter-to-waste period, at least two filters are mandatory, even for the smallest system. Two filters will also allow for periodic filter maintenance and for slow sand filter draining and scraping after each cycle.

6. The influent flow-splitting system used in the pilot plant of this study is an ideally simple system that would be appropriate to both rapid or slow sand filter plants for small installations. This arrangement (a) eliminates the possibility of sudden rate changes, (b) eliminates the possibility of negative head and consequent air binding, (c) eliminates the need for rate control equipment or head loss equipment, and (d) can be easily made fail-safe with a high water overflow-to-waste and a turbidity monitoring and automatic shut-down capability.

7. A good parallelism was evident for the three parameters of filtrate quality used in this study (turbidity, particle count, and total

coliform count). Thus a good job of continuous turbidity monitoring can give a good indication of particulate removal and should be an essential minimum of instrumentation for all plants, large or small, when a high degree of particle removal is essential on a continuous basis (for example, where Giardia lamblia may be present in the raw water).

The following general conclusions apply to direct, rapid filtration systems as applied to small water treatment systems treating high quality surface waters:

1. Declining-rate filtration did not produce better filtrate than constant-rate filtration in this application. Thus declining-rate filtration (which is more difficult to understand) should not be recommended for small systems. Influent flow splitting would be a superior system of operation.

2. A short period of flocculation of about 10 min should be provided in direct, rapid filtration. The flocculator should be provided with three or four compartments in series, a complete bypass to the filters, and bypasses at each compartment to allow flexibility in flocculator detention.

3. Chemical coagulants should be applied in direct, rapid filtration systems even when the raw waters are below the MCL of 1 NTU. Substantial numbers of particles can still be removed during such periods.

4. The research reported here demonstrated that the best direct filtration operation occurred during midwinter under ice cover with water temperatures of 2° C and with stable raw water quality. Cold water is therefore not an impairment to direct filtration.

5. Many existing conventional plants in northern climates could benefit by operating in the direct filtration mode during seasons of better raw water quality (e.g., in the winter during ice cover). Reductions in coagulant dosage, power, and maintenance costs could be achieved.

SECTION 4

RECOMMENDATIONS

1. Additional work would be desirable on the benefit of flocculation in direct filtration. Parallel studies of direct filtration with various flocculator detentions and various power input (G) should be conducted to supplement the work reported.

2. Future studies using coliform bacteria removal as a parameter of filter performance should be done with laboratory cultures rather than sewage as a source of bacteria, to provide more consistent coliform levels in the influent water.

3. Additional work would be desirable to confirm superiority of ferric salts over alum during periods of heavy algal blooms that elevate the pH above 8.

SECTION 5

LITERATURE REVIEW

WATERBORNE OUTBREAKS OF GIARDIASIS AND TREATMENT DEFICIENCIES

The need for adequate treatment of small community water supplies has been amply demonstrated by a number of recent outbreaks of waterborne giardiasis and other waterborne diseases of bacterial or undefined origin [21,22, 44,47,48].

Rome, New York experienced an outbreak of giardiasis in 1974 and 1975 with 4800 to 5300 estimated cases based on an epidemiological study [22,73]. Rome was served by a surface water source with chlorine and ammonia disinfection as the sole treatment. At the time of the outbreak, chlorine and ammonia were applied together to produce chloramine with a combined chlorine residual of 0.8 mg/L.

Camus, Washington had an outbreak of giardiasis in 1976 that affected 600 individuals in a population of 6000 [44]. Camus is served by both surface water and groundwater. The surface water was being treated by prechlorination in the transmission line to the plant with coagulant chemical addition at the plant followed immediately by filtration through multi-media pressure filters. *Giardia* cysts were isolated in the surface water entering the plant and *Giardia*-positive beavers were trapped in the watershed. The treated water reportedly met coliform and turbidity standards prior to and during the outbreak.

However, a number of deficiencies were found in the condition and operation of the filters and in the chemical pretreatment at Camus. Substantial loss of filter media had occurred and gravel mounding in the filters was reported. Both of these deficiencies could reduce the effectiveness of filtration. Three periods of loss of chlorine application occurred during the outbreak, but the onset of the outbreak preceded the first chlorination failure. The raw water turbidity during the period of concern rarely exceeded the 1 NTU standard for finished water, and the finished water continuously met the turbidity and bacteriological standards set by the Safe Drinking Water Act. It was concluded from this experience that "turbidity and coliform count alone are inadequate parameters on which to judge the biological quality of filter effluent" [44].

A similar outbreak involving a filtered water supply occurred in Berlin, New Hampshire, in which 750 cases of giardiasis occurred in 1977 [47]. The raw water was derived from two river sources from which *Giardia* cysts were recovered. Water from one source received pressure filtration without chemical pretreatment. Water from the other source was treated by a new plant providing chemical addition (alum, polymer, and sodium hydroxide), upflow clarification, and rapid sand filtration. Both plants provided post-chlorination. The pressure filters of the first plant were found to be in poor condition. Serious mounding of the surface of the medium, deep cracks in the medium along the walls, mud masses, and clogged areas were found. The air scour of one filter was broken so that all the air delivered during the backwash routine came out at one location, causing a deep depression in the surface of the filter medium.

With such filter deficiencies, it is little wonder that the filters did not provide good filtrate. Short circuiting through the cracks and depressions would occur, and the clogged areas would be inactive, forcing excessive flow rates through the unclogged areas. This represents a classic example of why some regulatory agencies do not permit pressure filters on surface water supplies [66].

Faulty construction of a common wall between the raw and treated water of the new plant permitted an estimated 3% of filter influent water to bypass the filters. This is another classic example of why most regulatory agencies do not allow common walls to exist between unfiltered and filtered water.

In spite of these deficiencies, routine bacterial samples collected from the distribution system prior to and during the outbreak did not violate coliform standards, leading the authors to conclude that "the coliform standard is not an acceptable indicator of safety where Giardia cysts are present" [47].

Other outbreaks of giardiasis have also occurred at Vail, Colorado in 1978 with 5000 cases [54]; at Bradford, Pennsylvania in 1979 with an estimated 2900 people affected [48]; and at Red Lodge, Montana in 1980 (personal communication, E. C. Lippy, Aug. 19, 1980). The outbreak at Vail was the result of inadequate filtration of surface water. The Bradford outbreak was the result of no treatment except chlorination for a surface supply, and the chlorination facilities were antiquated, had inadequate capacity, and the residual chlorine level was not properly maintained. The Red Lodge, Montana outbreak resulted from using only chlorination of a surface water with dosage levels below the cysticidal dose for Giardia cysts.

In spite of the weaknesses evidenced by the filters in the outbreaks described above, one would expect that granular deep-bed filters should do a good job in removing Giardia cysts if the filter were properly operated and maintained. The cysts are fairly large (about 8 to 12 µm by 7 to 10 µm [54]), and they exhibit a negative zeta potential of about -25 millivolts [54]. Thus, they should respond favorably to normal water treatment practices designed to remove negative particles commonly encountered in water.

It is apparent from the foregoing that the practice of providing only chlorination for protected-watershed, high-quality surface waters is not adequate to ensure protection of the public health. Furthermore, routine surveillance tests for coliform organisms and turbidity with satisfactory results do not give absolute assurance that a giardiasis outbreak cannot occur if Giardia cysts are present in the raw water.

It is also apparent from the foregoing that more than one barrier to disease transmission is needed to give added reliability to the system [54]. Furthermore, each barrier such as disinfection and filtration must be designed, operated, and maintained so that it serves its function effectively. The operation and maintenance requirements are especially difficult to ensure in the very small community. Thus, the treatment system should be simple, foolproof, and fail-safe to ensure the highest possible degree of public health protection.

SLOW SAND FILTRATION TECHNOLOGY

Many of the giardiasis outbreaks have occurred in upland watersheds utilizing low turbidity and low color water that presumably could be filtered effectively and economically using slow sand filters. The slow sand filter could serve as that important second barrier to prevent waterborne disease transmission. Because of this, it is important to determine what can be expected of future slow sand filtration by looking at past experiences.

Textbooks from the turn of the century and early 1900's are useful in determining general operating parameters and expected filtration performance. After reviewing a large number of early filtration textbooks, including Ellms [29], Hazen [36], Stein [79], Rideal [67], Ryan [68], Manual of British Water Engineering Practice [56], Flinn et al. [30], Hopkins [37], Baker [10], Horwood [38], Turneaure et al. [86], Norcom and Brown [62], and Babbitt and Doland [8], the following operating ranges and expected filtration efficiencies can be noted. Many of these parameters vary due to conditions unique to particular systems. The values presented here merely illustrate the ranges that existed in the late nineteenth and early twentieth centuries with the more common operating mode often noted.

The filter media generally consisted of a hard fine sand with an effective size ranging from 0.15 mm to 0.40 mm. The most common effective size was 0.30 mm. The uniformity coefficient of the fine sand was from 1.5 to 3.6. The majority of the sources reported a value of 2.0. The depth of this sand layer varied due to repeated scraping but had a suggested initial depth of 0.46-1.52 m (1.5 to 5.0 ft). Three feet was the predominant depth suggested by various authors. Below the fine filter sand was a graded gravel layer used to keep the filter media from entering the underdrain system. A three- or four-layer gradation was commonly used. Most systems incorporated a three-layer system ranging from 0.15-0.91 m (6-36 in.) deep. The United States used thinner gravel packs than those used in European countries. Graded gravel depths ranging from 0.46-0.61 m (18-24 in.) were the most common.

Flow rates through the media ranged from 0.04 m/h (1 mgad) to 0.4 m/h (10 mgad). Filtration rates above 0.12 m/h (3 mgad) were generally reserved for treatment facilities that incorporated some sort of pretreatment. Flow rates starting at 8 mgad were entirely served by pretreatment systems where the slow sand filter was used as a polishing step. The common flow rate for a low color and low turbidity water that had not received pretreatment was 0.08-0.12 m/h (2-3 mgad). Lower filtration rates were often used for the beginning or ripening stage of the filter run. A uniform rate of filtration was the main concern of many authors. A uniform rate was considered essential to avoid upsetting the schmutzdecke, which was considered to be responsible for the majority of the filter action.

The pressure needed to push the water through the media at the desired rate was supplied by the level of water lying above the sand. Available head loss ranged from 0.76-4.3 m (2.5-14 ft) of water in some cases. The majority of the plants in operation remained in the 0.9-1.5 m (3.0-5.0 ft) range to avoid having impurities driven too deeply into the filter bed.

The actual head loss through the media increased as the filter ripened and clogged. When the head loss had reached a predetermined amount, varying from 1.2-1.8 m (4.0-6.0 ft) above the initial head loss, the filter was scraped or raked depending on the system employed at the particular plant. Initial head losses ranged from 2.5-15 cm (1-6 in.).

When the filter reached maximum head loss the media surface was either scraped or raked. Scraping was used at the end of each run. Several raking cycles were often incorporated during the run to lengthen the run. When raking was part of the plant procedure, raking was utilized for one to five cycles before scraping was necessary. When raking was utilized during the run the raking cycles became increasingly shorter and the depth of penetration increased so that a deeper layer had to be removed when the filter was finally scraped. The final scraping may have required as much as six inches of media removal to assure a properly cleaned bed. When scraping was the only method of rejuvenation used, a layer 1-5 cm (3/8-2 in.) thick was removed. Generally the bed was clean if one inch of surface material was removed. The entire schmutzdecke had to be removed to avoid subsurface clogging at a later time.

The filter runs lasted from several days to more than three months. A great range of run lengths existed not only from plant to plant, but also from season to season in the same treatment facility. Since run lengths and raw water quality showed such a large fluctuation, the ripening periods reported also showed a great variation. A number of authors mentioned an 8-day-minimum ripening period. However, this may be unreasonable since some filter runs only last for a few days.

Bacteria removals reported by these early plants were often excellent. Removal rates of total bacteria generally ranged from 98 to 99%. Bacteria removal rates were higher when dealing with higher influent bacteria concentrations. However, even at lower influent bacteria levels, the removal rates only dropped to 85 or 90%.

Unlike bacteria results, poor color removals were reported by a number of the authors cited. Twenty-five percent color reductions were common for slow sand filters at the turn of the century.

The majority of the sources recommended slow sand filtration without pretreatment for waters that had less than 30 ppm turbidity and 20 ppm color (these units are as reported, although they are not the currently used units for turbidity and color). When raw water exceeded these limits it was suggested that pretreatment be used prior to filtration.

A variety of pretreatment methods was used. These included sedimentation, sedimentation with coagulant aids, preliminary filtration, and prechlorination. When the appropriate pretreatment methods were used it was possible to handle waters with turbidities in excess of 100 ppm.

Many slow sand filtration plants have demonstrated the ability to remove total bacteria and turbidity; however, few have demonstrated the ability to remove the pathogen *G. lamblia* associated with giardiasis. The lack of such studies has been due to the recent diagnosis of the problem and also due to the difficulty in keeping a supply of *G. lamblia* cysts available for a filtration study.

A recent paper by Bellamy, Silverman, and Hendricks [12] was the only study found which used *G. lamblia* cysts in a slow sand filtration study. The slow sand filtration pilot unit under study consisted of three slow sand filters operated in parallel at three different flow rates. In the study a batch feed tank was filled with lake water and then spiked with a known concentration of *Giardia* cysts. Influent and effluent sampling were separated by a 24-hour period to allow for the displacement of one filter volume of water. Cysts were not fed continuously during the study period. Spiking and sampling sessions lasted from 3 to 11 days. Cyst enumeration consisted of two steps, sampling and analysis.

Sampling consisted of concentrating a volume of water by passing it through a five-µm pore size, membrane filter. The filter was then washed, and the washings were collected for analysis.

Analysis consisted of two microscopic counting techniques. The first method involved floating the cysts onto a cover slip and then counting all the cysts. The second method made use of the micropipette technique. This technique involved reducing the sample to one mL by centrifugation, taking a 0.05 mL aliquot of the reduced sample, and counting the cysts in the aliquot.

Flow rates used in the study were 0.04 m/hr, 0.12 m/hr, and 0.40 m/hr. In addition to flow rate, temperature, cyst concentration, age of schmutzdecke, and the age of the sand column were varied to establish their relationships with the removal of *G. lamblia* cysts, bacteria, and particulate matter.

When the new sand was used in the study, filter efficiency was rather poor with mean removal rates of 99.15% and 84.00% for *Giardia* cyst and total coliform bacteria, respectively. This compared with mean removal rates of 100.0% (within detection limits) and 99.90% for *Giardia* cyst and total coliform removals, respectively, for a control filter with a mature sand layer. Further tests indicated the complete removal of cysts in the absence of a formed schmutzdecke after a six-inch-deep raking procedure. Additional tests showed that just after the filter was scraped, coliform removal rates of 99% (2 log) were achieved, as compared to 99.9% (3 log) removal rates for a filter with an established schmutzdecke. When the filter was totally resanded, the coliform removal rate was only 93%. These tests indicated that the steady improvement of cyst removal with time had little correlation with the removal of the schmutzdecke. This indicated the relative nonimportance of an established schmutzdecke. The authors concluded that a mature biopopulation throughout the filter was of primary importance.

During the study it was found that coliform removals were a better indicator of biological filter maturity than cyst removal rates. Therefore they suggested that bacteriological testing be used to measure filter performance rather than Giardia cyst levels. Temperature ranges from 5° C to 15° C, cyst concentrations from 50 to 5075 cysts per liter, and hydraulic loading rates from 0.04 m/hr to 0.40 m/hr were not found to affect cyst removal rates. However, the hydraulic loading rate did affect bacteria removal rates. Higher hydraulic loading rates produced lower bacteria removal rates. With a properly operated system it was concluded that slow sand filtration would be an effective form of water treatment that should be considered for all new systems, especially where a lower level of technology was desired [12].

Other slow sand treatment facilities supply information through records of bacteria and turbidity removals that may reflect on the ability of slow sand filters to remove Giardia cysts. Included here are several studies of pilot and full-scale plants that have demonstrated the ability of the slow sand filter.

A pilot plant study of slow rate filtration made by Fox et al. [32,50] demonstrated a highly efficient filter for the removal of turbidity and bacteria. The pilot plant filter contained clean builder's sand 0.76 m deep with an effective size of 0.17 mm and a uniformity coefficient of 2.1. The filter was operated at 0.12 m/hr while filtering two different water sources including low turbidity reservoir water (0.2 to 10 NTU), and high turbidity Ohio river water (0.4 to 23 NTU). The slow sand filter study lasted 800 days. After the ripening period, effluent turbidities were always below 1 NTU. As the filter aged, the effluent turbidities dropped below 0.35 NTU for the last half of the study. After the establishment of the filter biopopulation, that is, after the fourth scraping, a 50% increase in flow rate did not deteriorate the filter effluent water quality. Samples for total coliforms and standard plate counts were taken. Bacteria reductions of four to five log cycles were achieved continually after the initial ripening period. Effluent coliform densities were usually below 1/100 mL. Particle counts were also made using a Hiac Particle Size Analyzer. The slow sand filter demonstrated one to two log reductions for the particles in the 7 to 12 μm range. As the filter media matured, the removal rate increased. During the study, the drop in dissolved oxygen levels across the filter indicated aerobic biological activity in the upper portions of the filter. The mean influent and effluent dissolved oxygen levels were reported to be 7.85 mg/L and 7.33 mg/L, respectively. It was concluded from the study that slow sand filtration may be a good process for small systems not complying with the drinking water standards.

In another pilot study by Baumann, Willrich, and Ludwig [11], the efficiency of a slow sand filter with and without a prechlorination pretreatment was demonstrated. Filtering through 0.76 m (2.5 ft) of sand at a rate of 0.17 m/h with a constant head of 0.76 m (2.5 ft) of water and with a water temperature that ranged from 20°-34° C (67°-97° F), the filter without prechlorination removed 84.60% of the coliform bacteria and 90.38% of the total bacteria. In an identical slow sand filter that used prechlorination at an average free available chlorine residual, above the filter, of 8.8 mg/L, the reductions for coliform bacteria and total bacteria by the combined action of filtration and disinfection were 100.0 and 99.54%, respectively. In addition to this, the prechlorinated filter runs were longer and the penetration of particulates into the filter media was less. This study demonstrated the added advantages of combining a pretreatment step with slow sand filtration. It must be remembered that the raw water source in this study was river water. Even though the filter influent had been passed through a roughing filter to reduce turbidity, this raw water source may not have been ideally suited for slow sand filtration. Filter runs were short, and bacterial reductions without chlorine are much lower than reported in other studies with more appropriate raw waters.

Studies of actual community water treatment facilities give excellent insight into what can be achieved on a larger scale. Early data were often selected because operating data after 1909 were often influenced by chlorination.

From 1897 to 1898 a study was made for the city of Pittsburgh, Pennsylvania [45] to determine if slow sand filtration could be used to supply Pittsburgh with potable water. In this study two filters were operated to determine the bacterial efficiency of the process when treating Allegheny River water after a 24-hour sedimentation period. The filters consisted of 1.3 m (4.2 ft) of sand with an effective size of 0.27 mm and a uniformity coefficient ranging from 2.6 to 2.7. The media rested on a graded gravel pack. The filters were operated at a flow rate of 0.08 m/h (2 mgad) for less than a month at which time the flow rate was increased to 0.12 m/h (3 mgad) for the remaining 12 months of the study. The filters were operated until the head loss through the media reached 1.2 m (4 ft). At this time the filters were scraped and put back into service. At the end of the 13-month study the sand depths in the two filters were 0.88 m (2.9 ft) and 0.94 m (3.1 ft). Sedimentation preceded filtration.

During the first seven days of filter operation the average total bacteria removal was only 45 to 47%. After this ripening period the average monthly total bacteria removal rates for the 13-month study were all above 97%. For 6 of the 13 months the total bacteria removal rates were in excess of 99% [45].

According to Willcomb [89], MacHarg [55], and Baily et al. [9], early filtration at the Albany, New York treatment plant from 1899 to 1908 relied on slow sand filtration without pretreatment to treat Hudson River water. According to Willcomb [89], the early filters consisted of a 1.2 m (4 ft) deep sand layer with an effective size of 0.30 mm covering a 12-inch-deep gravel layer consisting of 3 gradations. The water above the filter provided

1.2 m (4 ft) of pressure head to force the raw water through the media at a rate of 0.12 m/h (3 mgad). During the first year of operation, total bacteria removals ranged from 94.8 to 99.7% while filtering Hudson river water with total bacteria levels ranging from 4,733 to 66,000 bacteria per mL [9]. Also according to Baily et al. [9], raw water turbidity was often 0.035 on the Platinum-wire standard with all turbidity removed upon filtration. The highest raw water turbidity recorded for the first year was 0.60. At this time the filters were delivering effluent water with a turbidity of 0.008 on the Platinum-wire Standard (an early turbidity standard of undetermined detail). Color removals were made on a platinum scale. Raw water color ranged from 0.50 to 0.60. Color removals ranged from 25 to 40%. MacHarg [55] considered the filters at Albany to be a great success in reducing typhoid deaths.

In England, the London filters demonstrated an equally impressive performance in treating Thames River water. Five different plants that drew water from the Thames river provided the filtered water supply of London. All five of these systems allowed some sedimentation during raw water storage and all reported bacteria removal rates higher than 99% [28]. Table 1 illustrates the details and total bacteria removal rates of each of the five treatment plants serving London on February 17, 1898.

TABLE 1. SLOW SAND FILTERS SERVING LONDON, ENGLAND IN 1898

Name of Waterworks	Depth of Gravel Layer (m)	Depth of Sand (m)	Flow Rate (m/h)	Percent Total Bacteria Removed
Chelsea	1.1	0.91-1.4	0.085	99.93
West Middlesex	0.30	0.76-0.84	0.061	99.70
Southwark	0.76	0.68-0.91	0.073	99.75-99.93
Grand Junction	0.76	0.61	0.073	99.10-99.84
Lambeth	1.2	0.76-0.91	0.081	99.90

Results similar to those reported were evident in many different community treatment facilities. The slow sand filters at Torresdale, Pennsylvania [63], in Berlin, Germany [33], and at Lawrence, Massachusetts [34], indicated similar results with total bacteria removal rates in the upper ninety percent range.

The abilities of slow sand filters for bacterial removal are well established and have been for years. It is evident that slow sand filtration is a viable alternative for the small community water system treating high quality water sources.

RAPID FILTRATION TECHNOLOGY

If rapid filtration of surface water supplies is used as the second barrier for public health protection, it should be done in full recognition of its strengths and weaknesses. For example, it is well known that the quality of the filtered water is poorer at the beginning of the filtration cycle and may also deteriorate near the end of the cycle [4,17]. Furthermore, any sudden increases in filtration rate on a dirty filter can cause breakthrough of deposited solids into the effluent [19,50].

The initial water quality degradation period has also been demonstrated in recent studies using *Giardia* cysts [54]. *Giardia muris* was used as a model for the human pathogen, *Giardia lamblia*. *G. muris* was spiked into a low-turbidity surface water, coagulated with alum alone or alum and cationic polymer, flocculated, and filtered through granular media filters. Initial cyst concentrations in the filtrate were from 10 to 25 times higher than those following the initial improvement period.

In conventional water treatment practice, the turbidity passage during the initial degradation and improvement period is small, averaged over the entire filter run. Because of this, the early practice of filtering-to-waste at the beginning of the filter run, to eliminate the turbidity carried through into the finished water, has been largely abandoned. It is evident that elimination of the "filter-to-waste" period may not be justifiable in the light of recent trends to use higher filtration rates in direct filtration applications. Also, it would be unacceptable where giardiasis is of concern because of the low infective dose for *Giardia* transmission [44] and the resistance of *Giardia* cysts to disinfection [43].

Evidence was presented in 1963 [19] showing the deleterious effect on the quality of filtered water when sudden rate increases are imposed on a dirty filter. The amount of material flushed through the filter was greater for sudden rate increases than for gradual changes. The amount of material released was greater for large increases than for small increases, but the amount was not affected by the duration of the maximum imposed rate. Different types of suspended solids encountered at different water plants exhibited different sensitivities to the rate increases. Similar observations have been reported in later papers [26,50,90]. These observations are important in any filtration decisions. Sudden rate increases on dirty filters should be avoided or minimized.

The recent studies by Logsdon et al. [50,54] also showed similar effects when the filtration rate was suddenly increased from 11 to 27 m/h. Turbidity in the effluent rose sharply and then rapidly declined. *G. muris* cyst concentrations followed the turbidity trends. "A four-fold increase in turbidity was accompanied by a twenty-five fold increase in cyst concentration in the filtered water" [54]. The same study demonstrated the detrimental effect of loss of coagulant feed and of extending the filter run into the period of terminal breakthrough. In both instances, large increases in cyst concentration were observed in the effluent. Thus, filtration systems which provide no effluent rate manipulation look very attractive.

DIRECT FILTRATION TECHNOLOGY

Direct filtration is the term applied to rapid filtration using coagulants in pretreatment but excluding sedimentation. All natural and added solids removal and storage occur in the filter media. Prefiltration treatment is to develop a filterable rather than settleable floc. The primary advantage of direct filtration is cost savings. Culp [23] indicates capital cost savings up to 30% and possible chemical cost savings of 10-30% due to less alum required for a filterable floc than for a settleable floc.

The direct filtration method of water treatment was first attempted in the early 1900's when the rapid sand filter was replacing the slow sand filter [23]. But the small media size (sand only) caused short filter runs. There has been a resurgence of interest in direct filtration for several reasons, among them: 1) the use of larger effective size anthracite coal in dual media filters; 2) the development of polymers which can be used in coagulation and can possibly reduce the floc volume; and 3) the passage of the Safe Drinking Water Act (SDWA) in 1974 establishing a 1 NTU MCL on finished water turbidity. This renewed interest in direct filtration has yielded papers on direct filtration including general overviews [20,23,49,50], research work and status in Canada [41,42], United States pilot plant and full-scale studies [58,61,82], and descriptions of existing or proposed direct filtration plants [75,81,83].

Appropriate Raw Waters for Direct Filtration

While direct filtration offers cost savings, it must not be applied to water with excessive turbidity, color, or algae because short filter cycles will be a problem. For this reason, considerable attention has been directed to define raw water quality criteria that are appropriate for direct filtration. The following paragraphs present some of those attempts.

Culp [23] indicated that direct filtration is viable if "1) the raw water turbidity and color are each less than 25 units; 2) the color is low and the maximum turbidity does not exceed 200 tu; or 3) the turbidity is low and the maximum color does not exceed 100 units." Culp also indicated that the diatom concentration should be less than 1000 asu/mL for direct filtration and recommended coarser anthracite if diatoms exceeded 100 asu/mL. McCormick and King [58] suggested a turbidity range of 0-10 NTU, a color range of 0-15 APHA color units and 0-1000 units/mL of algae (clump count). They suggested, however, that individual parameters higher than these could be tolerated when the other parameters were low.

A recent AWWA Committee report [20] suggested that color exceeding 30-40 Hazen color units (platinum-cobalt standard) or turbidity greater than

15 FTU on a continuous basis could be expected to cause short filter cycles. The report defined the following criteria for "perfect candidate" waters for direct filtration:

Color < 40 color units

Turbidity < 5 FTU

Algae < 2000 asu/mL

Iron < 0.3 mg/L

Manganese < 0.05 mg/L

Wagner and Hudson [87] evaluated waters for direct filtration by adding coagulant and flocculating in a jar test followed by filtration through Whatman No. 40 filter paper. Waters requiring more than 15 mg/L of alum to produce acceptable filtered water based on turbidity and color were doubtful candidates for direct filtration. Those requiring less than six or seven mg/L alum plus a small dose of polymer were considered excellent candidates for direct filtration. Those requiring doses between these levels were considered marginal candidates for direct filtration.

Those opinions were supportive of an informal opinion expressed in 1976 by personnel of the Water Research Center of Great Britain, who felt that alum requirements for treatment of upland, low-turbidity, colored waters should be less than 20 mg/L for direct filtration. This usually limited raw water color to about 40 color units.

Hutchison [41] reported the results of pilot scale studies of direct filtration of lake waters serving Ontario, Canada. He concluded that turbidity levels should be low enough to allow alum dosages of less than 12 mg/L on a continuous basis which would result in runs of 16-20 hours at 12 m/h (4.8 gpm/ft^2). Sporadic periods of turbidity requiring not more than 20 mg/L of alum could be tolerated on a short-term basis, but feeding of a nonionic polymer as a filter aid was then necessary to prevent terminal breakthrough.

Colored waters containing 25 to 30 color units in the raw Ontario waters would require alum dosages of more than 20 mg/L and result in short filter runs. However, the use of polymer as a primary coagulant to reduce or replace the alum might allow direct filtration to be applied to colored waters as high as 25 color units [41].

Diatom levels between 1000 and 2000 asu/mL required the use of coarser anthracite and more frequent use of polymer to prevent breakthrough. Anthracite with 1.5 mm es could handle diatoms to 2500 asu/mL and produce 12 h runs at 12 m/h (4.8 gpm/ft^2). However, diatom levels greater than 5000 asu/mL for prolonged periods would best be handled by other means such as sedimentation or microstraining [41].

From the foregoing literature, there has been considerable disagreement on the specific limits for turbidity, color, and algae. This is not surprising, considering the variability of coagulant dose required for a given level of these parameters. It would appear that the early estimates of Culp [23] are too liberal, while the later "perfect candidate" waters of the AWWA committee may be too conservative, particularly with regard to turbidity. There is more agreement with regard to the upper alum dose acceptable for direct filtration with 12 to 15 mg/L being a desired upper range. Higher dosage up to 20 mg/L can be tolerated if nonionic polymer is used as a filter aid to prevent terminal breakthrough of turbidity before reaching terminal head loss.

Chemical Pretreatment for Direct Filtration

An AWWA Committee survey [20] reported on 64 direct filtration plants. Eleven used metallic coagulants, either aluminum or iron salts, 12 used polymer only, and 13 used both polymer and metallic coagulant.

Some of the factors in the chemical choice were illustrated by the work of Hutchison [41]. Alum produced a low effluent turbidity over a pH range from 7 (the lowest used in the study) up to about pH 7.8, but turbidity deteriorated seriously above pH 8. In addition, the residual aluminum in the filtered water increased at higher pH.

The increase in residual aluminum with pH would be expected because the solubility of aluminum increases one log per pH unit above its minimum solubility pH of about 5.5, reaching about 0.05 mg/L at pH 7 and 0.5 mg/L at pH 8 expressed as Al (0.6 and 6 mg/L respectively as filter alum) [3]. Soluble aluminum passing through the filters can lead to after-precipitation of $Al(OH)_3$ in the distribution system if the pH is lowered by chlorination or other chemical changes.

Hutchison [41] also used ferric chloride in pilot scale and plant scale experiments. The ferric salts required about one-third the dosage of alum. Unfortunately, the form, molecular weight, and purity of the ferric chloride were not stated so the comparison of molar dosages cannot be made. If the chemical was anhydrous $FeCl_3$, the molar dosage would be about the same as the alum dosage. If it were crystalline $FeCl_3 \cdot 7H_2O$, the molar dosage would be only slightly lower for the iron salt. Nevertheless, the $FeCl_3$ was successful with no instance of terminal breakthrough by turbidity, even with a coarse anthracite with es of 1.55 mm. Furthermore, iron has a much lower solubility at typical water treatment pH than alum so the iron residual in the filtered water remained below 0.05 mg/L in all the tests.

Cationic polymer has become increasingly popular as a primary coagulant in direct filtration. It results in lower rate of head loss development and longer filter cycles for a given raw water [41,58]. This results because the polymer does not form a flocculant hydroxide precipitate to clog the bed as occurs with either aluminum or iron salts. The use of cationic polymer alone results in less tendency for terminal breakthrough of turbidity than the use of metallic coagulants, but it generally results in somewhat poorer filtrate quality [41,58].

Hutchison was not able to achieve a filtered water of <1 FTU using cationic polymer alone when the raw water turbidity was <5 FTU. Between a raw turbidity of 5 and 150 FTU, an effluent of 0.8 to 1.1 FTU was achieved using cationic polymer alone. This is not very impressive filtrate quality. For raw turbidity <5 FTU, it was necessary to use a small amount of alum (2 mg/L) in conjunction with the cationic polymer to meet a 1 FTU goal. The use of polymers is also attractive because they produce less sludge to be disposed of and sludge which may dewater more readily [20,51,90].

Cationic polymer alone has been shown effective in direct filtration for treatment of humic (colored) waters [27]. This study of a natural colored water in New York also showed that flocculation resulted in a lower rate of head loss development in direct filtration, and that pH had a small effect on filter performance over the range of 5.5 to 7.5. Three different commercial low molecular weight, poly quarternary amine, cationic polymers were equally effective (Betz 1190, Magnifloc 572C, and 573C). The required dosages, however, were quite high (8 to 15 mg/L for a raw water with 130-170 Pt-Co color units) to produce a filtered water with 5-15 color units.

In contrast, Fulton [35] reported that in the northeastern U.S. many colored waters have color characteristics which render polymers ineffective if used as a primary coagulant, and their effectiveness varies with seasonal changes. He reported that in this region, alum has been used with positive results to overcome these problems.

Because of the drawbacks of using a metallic coagulant or cationic polymer alone, there has been a growing trend to use combinations of the two, hoping that the strengths of one will offset the weaknesses of the other. For example, reduction of the alum dose should lower the rate of head loss development and reduce the tendency for terminal breakthrough. Addition of the cationic polymer will substitute for the reduced alum as a primary coagulant and it exhibits less tendency for terminal breakthrough. If color is present, the alum will work more effectively than the polymer in color coagulation.

Examples of the use of cationic polymer and metallic coagulant together are available [41,57,58,74,82,84]. Some typical dosage combinations reported to be effective are: (1) 2 mg/L alum and 2 mg/L Cat-Floc T for Owens River water, Los Angeles, California, which had a median turbidity of 2.8 NTU and a range from about 1 to 14 NTU [57]; (2) 3 mg/L alum and 0.25 mg/L Cat-Floc T for Deer Creek Reservoir water in Utah, which had an average raw water turbidity of 2.6 NTU and a range from 0.1 to 60 NTU [82]; (3) 10 to 12 mg/L alum and cationic polymer about one tenth the alum for Lake Mead water at Las Vegas, Nevada [61], a raw water with turbidity usually <1 NTU but algae from 1000 to 8000 per mL between April and December [75].

There is also some use of nonionic polymers for direct filtration. These are usually added to the filter influent to reduce the terminal breakthrough tendency [41]. The data available to the AWWA committee in preparing their report [20] also indicated some addition of nonionic polymers in the rapid mix or flocculation tank. In this case they may serve to strengthen the floc for better filtration but would not be serving as a primary coagulant.

The Impact of Flocculation on Direct Filtration

The growing interest in the use of direct filtration has raised the question of whether flocculation should be provided between rapid mixing and filtration. There have been a number of studies of this question leading to some conflicting observations.

Adin and Rebhun achieved good removal of clay using only cationic polymer as a primary coagulant in a laboratory study [1]. This led them to conclude that flocculation was not needed in direct filtration, although no parallel studies were made with flocculation.

Culp [23] summarized the data from an earlier nationwide survey of direct filtration plants [46] in which 8 of 17 plants had flocculation and 9 did not have flocculation. Culp concluded, without presenting experimental evidence, that "if a properly designed rapid mix is provided, then there is no reason to include flocculation in the direct filtration process."

In contrast, a number of pilot studies have demonstrated that a short period of flocculation can be beneficial to the filtrate quality and in reducing the rate of head loss development. However, flocculation also increases the tendency for terminal breakthrough of particulates.

Hutchison [41] reported on pilot scale comparisons of direct filtration with flocculation times of 4.5, 14.5 and 28.5 min (experimental time based on tracer tests) using dual media with 3 different anthracite sizes. The various times were studied in sequential rather than parallel runs, which is not the best research design. Increasing the flocculation time above 4.5 min increased the tendency for turbidity breakthrough; breakthrough occurred at lower head loss and shorter run times. Power input during flocculation was compared at velocity gradients (G) of 20, 100 and 300 s^{-1} at a detention of 14.5 min using the same three filter media. This was also done in sequential filter runs. The breakthrough was worse at 300 s^{-1}. Turbidity and head loss were nearly the same at 20 and 100 s^{-1}. Hutchison concluded that the probability of filter breakthrough in direct filtration is increased by: (1) increasing the es of the coal; (2) increasing the filtration rate; (3) increasing the flocculation gradient above 20 s^{-1}; (4) increasing the flocculation time to more than 10 minutes; and (5) decreasing the depth of media. No conclusion was offered on the effect of flocculation time on the quality of filtrate and probably none was justified because of the sequential arrangement of the filter runs. The effluent turbidity varied by only 0.03 FTU or less in any sequential comparison.

Treweek [84] compared flocculation times of 0, 2, 7, 15, 30, and 45 min at a G of 100 s^{-1} prior to direct filtration of Deer Creek reservoir water in Utah. Coagulants were 3 mg/L alum and 0.25 mg/L of Cat-Floc T. Water was flocculated in a 6-place gang stirrer (jar test apparatus), and filtered in batch experiments through 30 cm of fine sand with es of 0.5 mm in a 2.5 cm diameter column at a rate of 11.5 m/h. All six liters were filtered in sequence with samples taken after 2, 4, and 6 L had passed through the filter column; this would result in about a 60-min filter run. Because of the type of experiments, the various flocculation times were tested in series rather than in parallel. Results were compared based on turbidity and particle count analysis of the raw, flocculated, and filtered water. The results indicated a flocculation time of less than 7 min (G · t = 42000 at G = 100 s^{-1}) was not sufficient to produce aggregation of the singlet particles required for their most effective removal in the filter media. Increasing the flocculation time beyond 7 min resulted in larger and more visible floc in the stirrer but did not improve the quality of the filter effluent particle count or turbidity. Furthermore, a comparison was made between the direct filtration results noted above with a conventional complete treatment plant operating in parallel with the conclusion that the direct filtration and conventional processes produced comparable effluent particle count and turbidity results.

The principal weakness of the above study is that the tendency for terminal breakthrough and for head loss development could not be determined in these batch type direct filtration experiments.

McCormick and King [58] presented pilot scale direct filtration data collected at 5 cities in Virginia. They operated 3 filters with different media in parallel and compared performance with and without flocculation in sequential studies. Thus, the results may be clouded by changes in raw water from day to day or by minor changes in chemical feed rates. Flow from the rapid mix tank went either to a stirred flocculation tank with 28-min detention (G = 20 - 63 s^{-1}) or directly to a pumping reservoir with less than 10-min detention. Effluent from the flocculator, when in service, also flowed to the pumping reservoir. Thus, the results are also clouded by the flocculation that would occur in the pumping reservoir in service in all cases, whether the stirred flocculator was in service or not.

The authors stated that inclusion of flocculation in alum-coagulated trials reduced the rate of head loss development for a triple (mixed) media filter. They also concluded that the flocculation was not needed for the two dual-media filters which were studied; these filters used deeper anthracite of coarser grain size. However, a careful scrutiny of the data presented does not provide any convincing evidence to support these conclusions. Sequential runs were seldom unchanged in all respects except for the provision of flocculation. Changes in chemical dosage or type of chemical or raw water turbidity prevent valid comparisons in all but two cases. One case using alum (Run A-5 versus A-1) supports the first conclusion. The other case using cationic polymer (Run B-17 versus B-20) yields the opposite conclusion.

McBride [57] summarized direct filtration studies of Owens River water at Los Angeles using various flocculation schemes. A comparison of 20 min flocculation at G of 70, 175 and 420 s^{-1} showed no difference in filtrate turbidity and only slight differences in particle count. However, the unit filter run volume (UFRV) was lower at the lowest G, presumably due to a higher rate of head loss development.

A comparison was also made between 16 runs using direct filtration with 12 runs using direct, in-line filtration. Direct filtration (presumably 20 min flocculation at G = 70 s^{-1}) resulted in lower turbidity, slightly higher particle count and higher UFRV than in-line filtration. Again, the higher UFRV probably reflects a lower rate of head loss development. The observations led to the recommendation of 20 min of flocculation at G = 80 s^{-1}. A statement was made that the pilot studies showed that lower mixing energy inputs will perform satisfactorily, so a safety factor is included in the 80 s^{-1}.

Dharmarajah [25] made a laboratory pilot study of the removal of humic acid by direct filtration with and without flocculation. He used a flocculation time of 17.5 min, with G of 50 s^{-1}. The use of the flocculator did not improve the filtrate quality prior to breakthrough, but it did cause breakthrough to occur much sooner. Flocculation also reduced the rate of head loss development.

Sama [69] conducted a laboratory pilot study to evaluate the effect of flocculation time and velocity gradient on the length of the filter run in direct filtration. A suspension of 40 mg/L of bentonite clay and 20 mg/L of kaolinite clay was prepared in Chicago tap water, coagulated with Cat-Floc T and flocculated at various power inputs and for various durations, and then filtered. Filter runs were terminated either at a head loss accumulation of 228 cm (90 in.) or if the effluent turbidity exceeded 0.3 FTU. In one series of experiments, with a constant G of 90 sec^{-1} and 3 mg/L Cat-Floc T, flocculation time varied from 2.0 to 9.2 min. The length of filter runs was shown to increase as flocculation time increased from 2.0 to 6.0 min. Runs were all terminated by reaching the limiting head loss in this series rather than by breakthrough; thus the provision of flocculation lowered the rate of head loss development and lengthened the runs. In the same manner, using a constant flocculation time of 9.2 min and various G values (0, 25, 90, 275, 400, and 700 s^{-1}), the optimum run length was achieved at G = 90 s^{-1}. Again, all runs were terminated by reaching terminal head loss.

Hutchison and Foley [42] reported similar findings in pilot studies of direct filtration for Great Lakes water. Pilot scale studies were made with flocculation times of 8, 14.5 and 18 min at G = 20 s^{-1}. Alum dose was 15 mg/L and filtration rate was 17.6 m/h (7.2 gpm/ft^2). Filter runs were 10 h at 8 min flocculation with the run being terminated by reaching a terminal head loss of 2.4 m (8 ft). Runs decreased to 6.2 and 4.2 h at 14.5 and 18 min flocculation, which were terminated due to breakthrough at 1.7 m and 1.1 m (5.5 and 3.6 ft) head loss respectively. No experiments were done with flocculation times shorter than 8 min, but the authors reported severe after-flocculation problems at Port Elgin when the flocculation time was reduced to 3.5 min, but no such problems at 6.5 min (both presumably for cold winter conditions). ("After-flocculation" refers to the passage of alum through the filters and precipitation of $Al(OH)_3$ after leaving the filters in the clear well or distribution system.)

Stephenson [80] evaluated the ability of a direct filtration system to remove added kaolinite and aluminosilicate at flocculation times of 0, 5, and 10 min with a constant velocity gradient of 50 s^{-1}. He found that 5 min and

10 min flocculation time resulted in about equal head loss but both had lower head loss than at 0 min. Filtrate quality was the same for all 3 flocculation times.

Sweeney and Prendiville [81] reported on pilot experiments in Springfield, Massachusetts and stated that the results indicated a flocculation time of 30 min may be needed during periods of cold weather. However, no supporting data were presented.

Monscvitz et al. [61] reported on pilot studies on the effect of flocculation on direct filtration of Lake Mead water. For this low turbidity water with substantial algae populations, they found 20 to 30 min of flocculation at $G = 20\ s^{-1}$ resulted in minimum alum requirements. This conclusion was based on zeta potential adjustment and process performance indicated by filter effluent turbidity, filter run duration and head loss profiles. Provision of flocculation resulted in longer filter runs, more consistent plankton removal, and eliminated the need to reduce filtration rates when powdered activated carbon was being used for taste and odor control. Plankton removals greater than 90% were achieved by flocculation and filtration. As a result of these studies, the full-scale plant was designed with 30 min total detention in tapered flocculation basins with 4 cells in series. The G is adjustable from 75 to 25 s^{-1} in the first cell and is tapered to 30 to 10 s^{-1} in the fourth cell.

From the foregoing experimental studies, it seems well established that provision of some flocculation time reduces the rate of head loss development in direct filtration, but it also results in earlier breakthrough of turbidity. As flocculation detention times are increased, these trends are accentuated and no clear experimental evidence exists to justify times greater than 10 min, except for the Lake Mead study [61].

The earlier breakthrough of turbidity may actually result in shorter filter runs unless a filter aid polymer is used to retard terminal breakthrough. The filter aid polymer may partially or completely negate the head loss benefit resulting from flocculation.

The benefit of flocculation to the filtrate quality before terminal breakthrough has not been so clearly demonstrated by the experimental data reviewed. Only four studies contained such information [51,54,61,84]. Furthermore, no data were found that clearly showed the impact of flocculation on the initial improvement period of the filter run.

Filter Media for Direct Filtration

The choice of filter media for direct filtration is influenced by many factors. Since the entire solids load, both natural solids plus the solids from chemical addition and precipitation, must be stored within the filter media, the media should facilitate penetration of solids and removal over a substantial depth of the media. The most popular media in U.S. practice is dual media of coarser-grained anthracite on top of finer-grained sand. Other variations include triple media in which a finer layer of high-density media (garnet or ilmenite) is placed below the sand; or deep beds of coarse media, either sand or coal.

The performance of the media is affected by the other filtration variables discussed earlier such as chemical coagulant used and whether flocculation is provided. If flocculation is provided, the time and power input (G) affect the filter performance. Long flocculation times, or higher-power intensities, or both, favor deeper penetration of solids and may even hasten terminal breakthrough. Overdose of alum or inappropriate pH will cause deeper penetration and increase probability of terminal breakthrough. These penetration and breakthrough trends can be offset by the use of finer media or by the use of a nonionic filter aid polymer.

In reviewing the literature on filter media, it must be kept in mind that these other variables are influencing filter performance and media recommendations.

Hutchison and Foley [42] studied the effect of the coal size in dual media filters on effluent turbidity and head loss in direct filtration of Great Lakes waters in Ontario, Canada. A constant coal layer depth of 46 cm (18 in.), sand depth of 31 cm (12 in.) and a filtration rate of 11.7 m/h (4.8 gpm/ft^2) were maintained. They varied the effective size of the coal between 1.2 and 2.0 mm, with uniformity coefficients between 1.4 and 1.45, observing that the run time to reach terminal head loss of 1.2 m (4 ft) varied between 3.2 and 4.7 h for 1.2 mm coal and between 16 and 18.9 h for a 2.0 mm coal. They also reported that with 1.2 mm coal, 94% of the head loss (that is, floc storage) took place in the coal layer, but with 2.0 mm coal only 36% of the head loss occurred in the coal. Hutchison and Foley concluded that a coarse-coal layer with an effective size greater than 1.5 mm could handle diatoms up to 2500 asu/mL with filter runs exceeding 12 h at 11.7 m/h (4.8 gpm/ft^2). Furthermore, when using alum plus polymer when needed, a dual media consisting of 46 cm and 31 cm coal and sand depth, respectively, will produce a high-quality effluent with turbidity less than 0.3 FTU. These investigators asserted that if diatoms were not present, the best coal size for overall filter performance would be in the range of 1.0 to 1.1 mm es.

The effect of media depth on filter performance was also investigated by the same authors using 2.0 mm es coal with coal depths of 46 and 30 cm (18 and 12 in.). Increased water production from 60-100% per filter run was reported to be achieved when the 46 cm coal layer was selected. Terminal breakthrough was prevented by use of polyelectrolyte. Hutchison and Foley explained that such increase in run length was due to more floc deposited in the coal layer for the deeper bed, rather than in the intermixing zone as occurred for the shallower bed. Due to the wide size range between the coal and the sand, about 20-25 cm (8-10 in.) of intermixing occurred at the interface, and only about 10 cm (4 in.) of unmixed coal remained at the top.

McCormick and King [58] carried out experiments on direct filtration using coal of three different effective sizes: 1.05, 1.3, and 1.7 mm. The raw waters and chemical treatment were presented earlier. The coal depth used by these investigators was 51 cm (20 in.) over a sand depth of 25 cm

(10 in.) of 0.45 mm es. They reported that for the largest coal, the runs were terminated due to early breakthrough. The head loss accumulation was more rapid in the filter with the finest coal, that is, 1.05 mm. The 1.3 mm coal was reported to store more than 90% of the solids and in this case, the sand was used primarily to polish the filtrate, providing longer filter runs than either the 1.05 and 1.7 mm coals. They considered this dual media (51 cm of 1.3 mm es coal over 25 cm of 0.45 mm es sand) to be the most effective design for the 5 waters which they treated.

Tate et al. [82] conducted experimental work on direct filtration of Deer Creek Reservoir water in Utah using three different filter media to evaluate their effectiveness in removing particulate matter. The chemical treatment of this water was presented earlier. The three media are described in Table 2. Tate et al. reported that the highest head loss accumulation was observed to occur within the filter using anthracite and garnet, while the lowest accumulation was observed within the filter using anthracite and sand. They indicated that the turbidity results from the three filters did not show any significant difference in a practical sense; final turbidities between 0.18 and 0.11 tu were achieved. The turbidity of the raw water was reported to vary in the range of 0.57 to 0.72 tu. The particle size distribution within the finished water from each filter was observed to be virtually identical. Removal of particles with diameters from 2.5 to 150 μ was also observed to be greater than 99%.

TABLE 2. FILTER MEDIA STUDIED BY TATE ET AL. [82]

Filter	Media	Depth cm	Depth in.	Effective Size mm	Uniformity Coefficient
1	sand	25	10	0.5	<1.35
	anthracite	51	20	1.1	<1.35
2	garnet	11	4.5	0.26	1.73
	sand	23	9	0.38	1.47
	anthracite	41.9	16.5	0.88	1.36
3	garnet	15	6	0.21	1.33
	anthracite	61	24	1.3	1.42

Monscvitz et al. [61] compared the performance of two dual media at Lake Mead. The chemical treatment and flocculation provisions have been discussed earlier. The two dual media consisted of 51 cm (20 in.) of coal on top of 25 cm (10 in.) of sand. Sand was 0.45 mm es, uc < 1.5. Anthracites were 0.95 mm es initially and later changed to 0.7 mm es, uc < 1.5 in both cases. It was concluded that the dual media with 0.95 mm es produced more uniform head loss across the anthracite-sand media and was as efficient in turbidity and plankton removal.

McBride et al. [57] presented direct filtration results for four media types in the treatment of Owens River water at Los Angeles. The chemical treatment and flocculation studies were presented earlier. The most interesting results compare a dual media with 51 cm (20 in.) anthracite (1.1 mm es, uc = 1.24) over 25 cm (10 in.) of sand (0.51 mm es, uc = 1.20) with a coarse media filter with 2.4 m (96 in.) of sand (2.1 mm es, uc = 1.48). The operating results at 14.7 m/h (6 gpm/ft^2) indicated no statistically significant difference in the turbidity or particle count of the effluent; but the coarse sand filter had a 50% greater water production per run than the dual media. As a result of these and other studies at higher filtration rates, the deep coarse sand bed was recommended for the plant design at a filtration rate of 22 m/h (9 gpm/ft^2). Subsequent work using preozonation rather than prechlorination has resulted in a recommended filtration rate of 33 m/h (13.5 gpm/ft^2) and a 1.8 m (6 ft) deep bed of anthracite 1.4-1.6 mm es, uc < 1.5. The proposed Los Angeles plant is probably the most unconventional proposal to date in terms of pretreatment, filter media type and filtration rate.

Miscellaneous Direct Filtration Considerations

Yapijakis [[illegible]0] concluded that adding 0.05-0.1 mg/L nonionic polymer during the first half of the filter backwash reduced the initial turbidity breakthrough peak and period. Also, the settling and thickening rate of the backwash solids was enhanced. Dosing the last part of the backwash water with polymer to condition the filter has also been suggested [20].

Culp [23] provided some direct filtration design criteria. Hydraulic jumps and Parshall flumes for rapid mix have given good field experience. The usual filtration design rate is 12.45 m/h (5 gpm/ft^2). Filter influent and effluent piping should be designed for 24.9 m/h (10 gpm/ft^2). Surface or air wash should be included along with adequate water backwash rate. The effluent turbidity should be monitored on each filter and a coagulant control pilot filter should be used to optimize coagulant dosage in the full-scale plant. A nonionic or slightly anionic polymer filter aid (0.05-0.5 mg/L) should be used, with excess polymer added during initial filter operation.

The AWWA Direct Filtration Committee report contained other useful information regarding direct filtration [20]. During the design period, raw water quality records, climatological data, watershed control, and regulations should be evaluated. The suggested filter run length is 10-20 h. An anthracite size of 1.1 mm es was found to be optimum for dual media filters. Anthracite less than 0.8 mm es gave shorter runs and coal larger than 1.1 mm es required careful operation and polymer addition to prevent breakthrough. The importance of good rapid mixing of chemicals was emphasized. All plants should monitor turbidity, temperature, pH, bacterial content of the raw water, turbidity of each filter effluent and above the coal-sand interface, and the residual aluminum. Logsdon [49] and Logsdon et al. [54] stressed that adequate pretreatment even of clear waters is essential to ensure effective pathogen removal.

Hutchison [41] reported that the early direct filtration plants in Ontario, Canada were run at 6 m/h (2.4 gpm/ft^2) but in 1967, they were allowed to increase to 12 m/h (4.8 gpm/ft^2).

The above recommendations for filtration rates about 12 m/h (4.8 gpm/ft^2) seem rather conservative when compared with the latest Los Angeles recommendation of 33 m/h (13.5 gpm/ft^2) mentioned in the prior section. However, one should not assume such high rates are appropriate everywhere, even if the media and pretreatment were duplicated. Certainly, for smaller communities where operating skill or attention may be lacking, such high rates are not appropriate. Even 12 m/h (4.8 gpm/ft^2) may be too high in such cases. Lower rates would result in longer runs during troublesome raw water periods, and easier operation for the unskilled operator.

CONSTANT-RATE VERSUS DECLINING-RATE FILTRATION

One phase of the research reported herein involves an experimental comparison of constant-rate filtration and declining-rate filtration. Therefore a review of the literature on these methods is appropriate. Some prior knowledge of the subject is assumed or can be obtained from current textbooks [72,88]. These sources contain good illustrations and discussions of principles and operating characteristics of these two systems.

Constant-Rate Filtration

Constant-rate is a term applied to filters which have constant water throughput. The initial design of constant-rate (rapid sand) filtration was by Fuller in the 1890s as cited by Sanks [72]. This design was applied to clay-bearing waters in the United States which the English slow sand filter could not effectively treat. Constant-rate filtration (CRF) was synonymous with rapid sand filtration from the early 1900s until the alternate declining-rate filtration (DRF) method became established in the 1950s. Rapid sand (less than 7.5-10 m/h (3-4 gpm/ft^2)) and high rate (flows greater than rapid sand) filtration refer to classes of filter flow rates, whereas constant-rate and declining-rate refer to filter control methods.

Arboleda [6] indicated that when rapid sand filters were first introduced it "was thought that by keeping filtration velocity more or less constant during the runs a better effluent could be produced and an easier operation of the treatment-plant hydraulics could be achieved." Degremont [24] agreed with this statement but said that the "best control systems are...controllers that are simple to maintain and adjust...that operate without hunting...." Cleasby et al. [19] stated that there is a delicate balance between deposition and dislodging forces in the filter media. The effect of rate disturbances on effluent quality is a function of the magnitude of the disturbance and the time required for the disturbance. Greater flow increases and quicker disturbances increase the peak effluent concentration resulting from the disturbance.

Constant-rate filtration can be divided into three types: 1) constant water level, 2) constant rate, and 3) influent flow splitting. Only if the total plant influent is kept constant are these filtration types true constant rate. Variations in total plant flow will cause variations in filter flow. The constant water level and constant-rate control systems are essentially the same except that the water level can vary in the constant-rate system but is maintained constant in the constant water level system. Both are effluent-controlled systems. The constant-rate system requires manual adjustment for variations in influent flow, while the constant water level system adjusts the effluent control valve in response to flow variations to maintain a constant water level. The effluent constant rate controller consists of a rate-sensing device, a rate-setting device, and a control valve. There are two disadvantages of these two systems of mechanical effluent control [15]: 1) abrupt surges in influent flow and control valve 'hunting' have a detrimental effect on effluent water quality, and 2) the control devices are costly to install and maintain.

An influent flow splitting system contains an influent weir at the top of each filter box, which splits the flow equally to each filter (influent controlled). This system has no effluent controller; the level changes gradually during variations in influent flow or backwashing, and the water level is an indication of the head loss in each filter. An effluent weir must be provided with an overflow elevation above the filter media in order to avoid filter dewatering and negative head in the filter. Because of this weir, additional filter box depth is required [15,72].

Declining-rate filtration is presented in the following section. Comparisons of constant-rate and declining-rate control are also provided in the Declining-Rate Filtration section.

Declining-Rate Filtration

Declining-rate filtration is a control method that is distinctly different from the constant-rate methods previously described. True declining-rate filtration is defined as constant total head loss across the media [15]. However, in the practical application of declining-rate filtration (DRF), the rate is constant in each filter between backwashes. The flow rate declines in stepwise fashion through a run and therefore is sometimes called variable declining-rate filtration [15]. In a declining-rate filter plant with several filters, the water level in all filters is about the same due to a common influent header located below the operating low water level. This free communication between filters results in different flow rates through each filter depending on the extent of clogging. Degremont [24] describes the range of filter flow as +/- m% of the mean filter flow where the clean filter is at (1 + m/100)Q/N (Q/N = total flow divided by the number of filters), the dirtiest filter is at (1 - m/100)Q/N), and m is in the range of 20 to 40%.

The variations on DRF are unrestricted declining flow rate [6], and influent and effluent restricted DRF [6,40]. In unrestricted declining flow rate, no attempt is made to restrict the maximum rate through the cleanest filter. In influent restricted DRF, a fixed inlet constriction is used to limit the maximum applied flows. This constriction is not variable and is only useful in ensuring that the peak filter (clean filter) flow does not exceed a certain rate. In effluent restricted DRF, a fixed restriction is placed in the effluent pipe to limit the maximum flow through a clean filter. Cleasby and DiBernardo [18] limited the initial clean filter flow to 1.5 times the average rate.

Hudson [40] indicates that the first DRF system known to him was constructed in Albany, New York, in 1899. The system did not become popular and was not seen again in a larger plant until 1949 at the Howard Bend Plant in St. Louis. Subsequently, the DRF system received increased attention in technical papers and plant design. The majority of the United States designs have utilized effluent restriction, with the first influent restricted plant being constructed at Cali, Columbia, in 1978 [40].

The 1980 American Water Works Committee Report on Direct Filtration [20] indicated that the majority of direct filtration plants in the United States are constant-rate rather than declining-rate, that pilot plant results indicated little quality advantage of the DRF over the CRF, and that there was a concern of high initial rates in declining-rate filter operation. Taiwan and many South American countries have turned to DRF as a filtration system that does not require capital and maintenance costs for control equipment [6,15].

Discussion of declining-rate filtration in the technical literature began after constant-rate rapid sand and high-rate filters were common and accepted filter operation techniques. Therefore, DRF is usually discussed and studied as it compares to CRF [6,7,14,15,26,39,40,42, and 72].

Some of the most commonly discussed advantages and disadvantages of DRF are listed as follows:

Advantages:

1. DRF avoids random fluctuations in flow rate caused by rate controllers [6,40,72]. Flow variations cause a deterioration in effluent water quality [19].

2. DRF saves on initial and maintenance costs of control equipment [6,7,14,72].

3. Filtered water quality is better [14,15,39].

4. Diurnal rate variations occur gradually and smoothly without automatic control equipment [72].

5. For waters that show a breakthrough pattern, DRF gives better filtrate quality [72].

6. There is a lower floc shearing force in DRF than in CRF [42].

7. There is less available head required [15,18,26,39,40,72].

8. It is speculated that additional floc deposition areas within the filter are used as the flow rate decreases and therefore the filter runs are longer [42].

9. The water level rise during backwash is smaller because the dirtiest filter taken off-line for backwashing is 20-40% below average flow compared to average flow on the CRF [40].

Disadvantages:

1. There is uncertainty about how to calculate the total available head loss required [72]. (This has been explained by Cleasby and DiBernardo [18].)

2. High initial flow rates through clean filters may cause poor water from the cleanest filter at the beginning of the filter run [20]. (This has been explained by Arboleda [6] and Cleasby and DiBernardo [18].)

3. There is no advantage over constant-rate filtration [7,42]. (The papers cited were not well-controlled comparisons and did not provide all of the experimental evidence.)

The types of research conducted on DRF include conversion of existing plant scale facilities to DRF [7,16,39], examinations of plant scale DRF [6,15], pilot scale comparisons of DRF and CRF [6,26] and secondary investigations of DRF as a sideline to the primary research effort [42].

The classic temporary conversion of filtration to declining-rate was done by Baylis at the South District Filtration Plant in Chicago "with almost no decrease in the quality of filtered water" [7]. Baylis compared CRF in Gallery 1 with DRF in Gallery 2 by the membrane filter and cotton plug residue methods. Galleries 3 and 4 (CRF) were also included initially but then dropped from the comparison. Gallery 2 (DRF) was outperformed by Gallery 1 (CRF) but Gallery 2 outperformed Galleries 3 and 4 (CRF) in solids removal. Baylis concluded from his work that "unless subsequent work shows that there is a slight improvement in the quality of the water in most plants...that plants should be constructed with rate control." The declining-rate filters, however, were operated at higher rates and provided greater water production than the constant-rate filters. Arboleda [6] and Hudson [39] commented on the Baylis study in later works on the fact that the DRF run lengths were 50% longer than those on the CRF. Also, the representative flow and head loss charts included in the Baylis paper indicate very smooth variations for the DRF but very erratic curves for the CRF.

A second temporary conversion of plant scale facilities to declining-rate control was done by Hudson [39] at the 10 mgd Wyandotte, Michigan plant. The research was designed for and evaluated on the basis of a breakthrough index. Hudson concluded that DRF yields higher quality water and longer filter runs. Regarding permanent full scale conversions to DRF, Cleasby [16] offers suggestions on conversions of existing effluent-controlled CRF plants to DRF. The prerequisites for such a conversion were given along with some alternatives for controlling the total filter head loss.

Arboleda [6] used examples of plants in Antofagasta, Chile; Cochabamba, Bolivia; and Medellin, Colombia, to illustrate his explanation of filter hydraulic control. Arboleda concludes that "declining-rate filtration is more logical than constant-rate filtration, because it is inconsistent to force the filter bed to work at the end of the run, when it is clogged, with the same velocity as at the beginning of the run, when it is clean."

The only pilot scale research with the primary objective of making a CRF and DRF comparison was the study by DiBernardo and Cleasby [26]. They used a single constant-rate filter and 4 declining-rate filters to compare head loss, filter run time, and effluent water quality characteristics. The research conclusions included: 1) "substantially better" average DRF effluent turbidity; 2) longer DRF run length when both filter units were operated at the same filter media head loss increase; and 3) worse backwash and other rate variations on the CRF than those on the DRF. The research was conducted at 7.33, 12.22, and 17.11 m/h (2.94, 4.91, 6.87 gpm/ft^2, respectively).

Arboleda [6] constructed a 4-filter DRF pilot unit and reported on individual filter flow rates during backwash. He illustrated the transition through the backwash operation of a filter from the dirtiest (lowest flow) position to the cleanest (highest flow) position. Gregory and Yadav compared pilot scale DRF and CRF in England [personal communication, 1979]. Two surface water supplies were coagulated and settled prior to filtration. Although clean DRF flow was not restricted (5 times average flow rate), their declining-rate filter gave better effluent quality and longer filter runs than their constant-rate filter.

The secondary research effort in comparing the two filter control methods was by Hutchison and Foley [42]. They concluded that DRF had little advantage over CRF for the flow rates examined because the length of filter runs, effluent quality and floc distribution were similar at 12.45 m/h (5 gpm/ft^2). However, they neglected to point out the relative head loss increases in the filters. At 11.95 m/h (4.8 gpm/ft^2) on the CRF and 12.70 m/h (5.1 gpm/ft^2) average, on the DRF, the head loss increases were 0.073 m/h (0.24 ft/h) and 0.070 m/h (0.23 ft/h), respectively. At 22.2 m/h (8.9 gpm/ft^2), average, on both CRF and DRF, the head loss increases were 0.36 m/h (1.14 ft/h) and 0.20 m/h (0.67 ft/h), respectively. No influent or effluent turbidity data were given.

SECTION 6

EQUIPMENT AND MATERIAL

PILOT PLANT

The filtration plant was located on the east side of the southeast cell of Hallett's Quarry north of the Ames city limits on Highway 69. A 16-foot square wood frame building was set on a gravel base for drainage. Except for the raw water pump and the raw water influent and effluent drain lines, all of the equipment was inside the building. Figures 1, 2, and 3 are schematic presentations of the filtration equipment used during the three phases of the study; Fig. 1 represents Phase I during the slow sand and rapid filtration comparison, Fig. 2 represents Phase II during the declining-rate filtration (DRF) versus constant-rate filtration (CRF) comparison, and Fig. 3 represents Phase III during the direct filtration comparison with and without flocculation. Phase I began in Oct. 1981 and ended in Nov. 1982. Phase II began in June 1982 until Oct. 1982 and thus Phases I and II operated in parallel. Phase III operated from April 1983 through mid-July 1983.

Raw water was pumped to a splitter box mounted above the filters. The flow to be filtered was selected by the size of orifice delivering flow to each filter. The orifices were operated under a constant head created by an overflow weir that delivered excess flow to waste. Figure 4 presents the vertical arrangement of the filters and piezometer head loss scales for the CRF and the DRF. The arrangement shown for the CRF was the same for all three phases of the work. In Phase III, a second identical CRF was added as shown in Fig. 3.

Figure 5a presents the details of the influent splitter box used in all phases of the study. Figure 5b presents the details of the effluent collection trough and baffled mixing channel used to blend the effluents of the four DRFs before withdrawing the sample to the turbidimeter. Figure 6 presents the details of the flocculation tank and the flocculator paddles used in Phase III of the study. Figures 7, 8, and 9 are selected pictures of the research site and filtration equipment. Note that the doors into the shelter faced east and the high windows faced south.

The shelter, filters, influent flow splitter box, backwash water supply tank, and declining-rate filter effluent trough were all constructed by the Iowa State University Engineering Research Institute Shop.

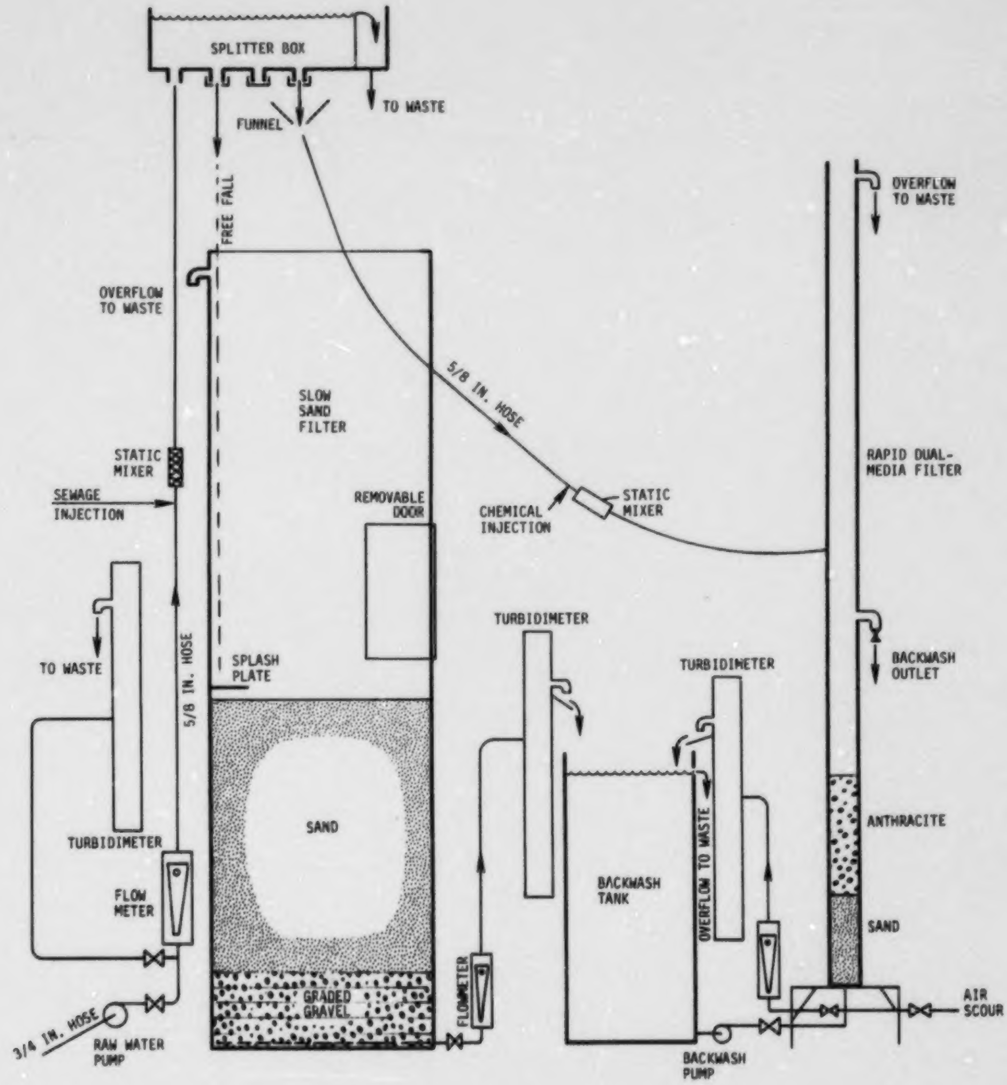

SPLITTER BOX
TO WASTE
FUNNEL
FREE FALL
OVERFLOW TO WASTE
OVERFLOW TO WASTE
SLOW SAND FILTER
5/8 IN. HOSE
STATIC MIXER
SEWAGE INJECTION
REMOVABLE DOOR
CHEMICAL INJECTION
STATIC MIXER
RAPID DUAL-MEDIA FILTER
TO WASTE
5/8 IN. HOSE
SPLASH PLATE
TURBIDIMETER
TURBIDIMETER
BACKWASH OUTLET
TURBIDIMETER
SAND
FLOW METER
BACKWASH TANK
OVERFLOW TO WASTE
ANTHRACITE
SAND
GRADED GRAVEL
FLOWMETER
AIR SCOUR
3/4 IN. HOSE
RAW WATER PUMP
BACKWASH PUMP

Fig. 1. Pilot plant schematic showing slow sand filter and rapid, dual-media filter used in Phase I.

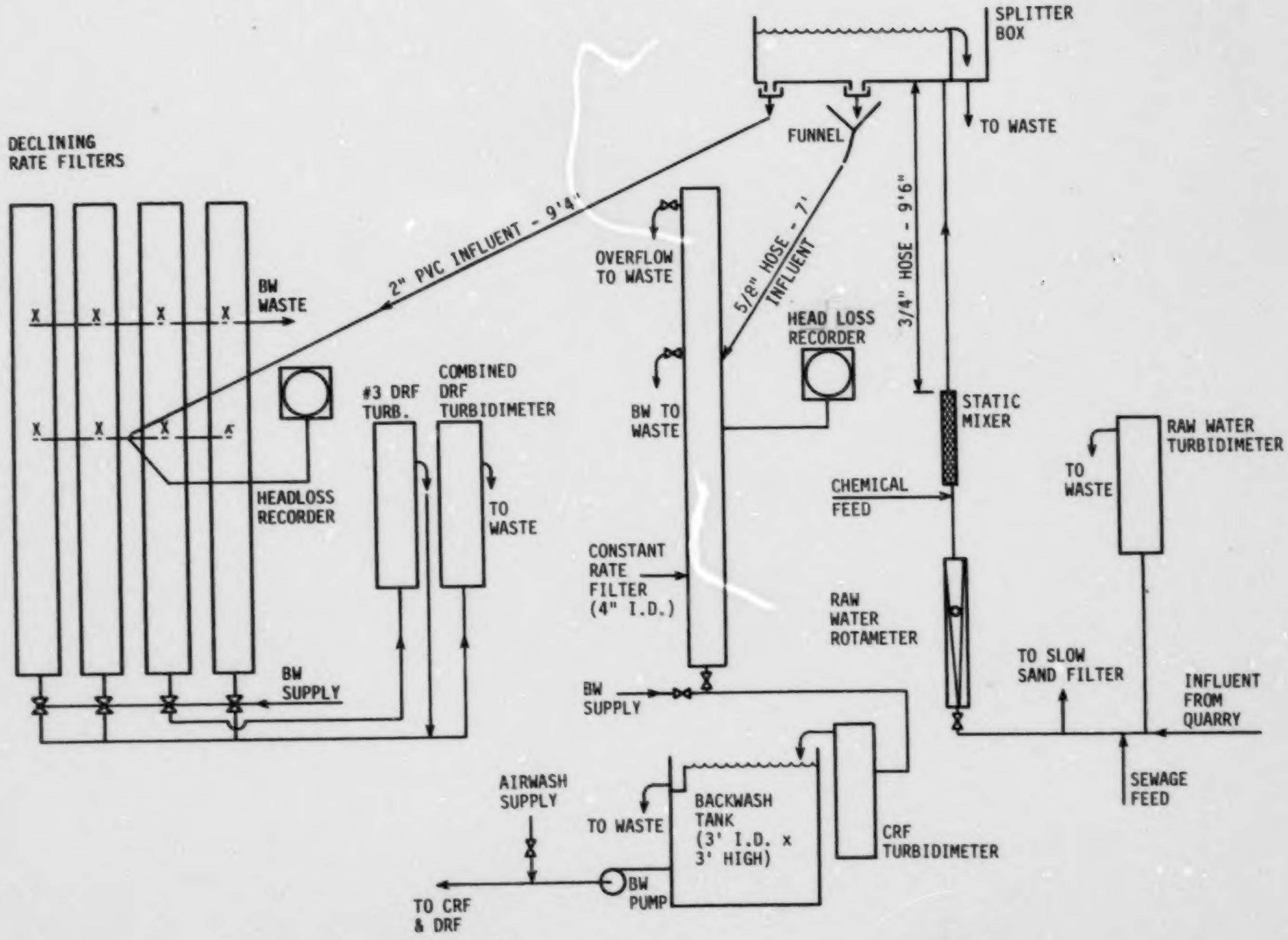

SPLITTER BOX
TO WASTE
FUNNEL
DECLINING RATE FILTERS
2" PVC INFLUENT - 9'4"
5/8" HOSE - 7' INFLUENT
3/4" HOSE - 9'6"
OVERFLOW TO WASTE
BW WASTE
HEAD LOSS RECORDER
STATIC MIXER
RAW WATER TURBIDIMETER
TO WASTE
#3 DRF TURB.
COMBINED DRF TURBIDIMETER
BW TO WASTE
HEADLOSS RECORDER
TO WASTE
CHEMICAL FEED
CONSTANT RATE FILTER (4" I.D.)
RAW WATER ROTAMETER
TO SLOW SAND FILTER
INFLUENT FROM QUARRY
BW SUPPLY
BW SUPPLY
SEWAGE FEED
AIRWASH SUPPLY
TO WASTE
BACKWASH TANK (3' I.D. x 3' HIGH)
CRF TURBIDIMETER
BW PUMP
TO CRF & DRF

Fig. 2. Pilot plant schematic showing four declining-rate filters and one constant-rate filter used in Phase II.

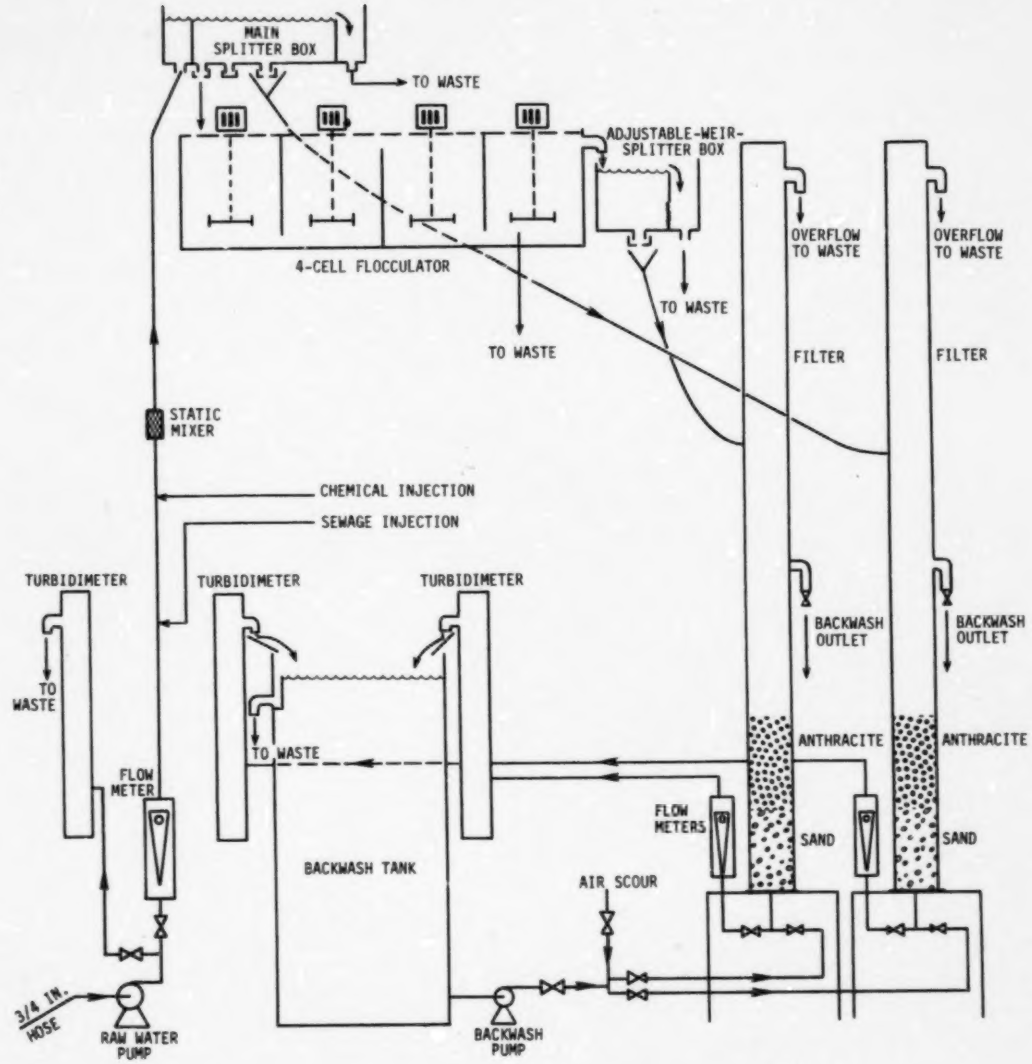
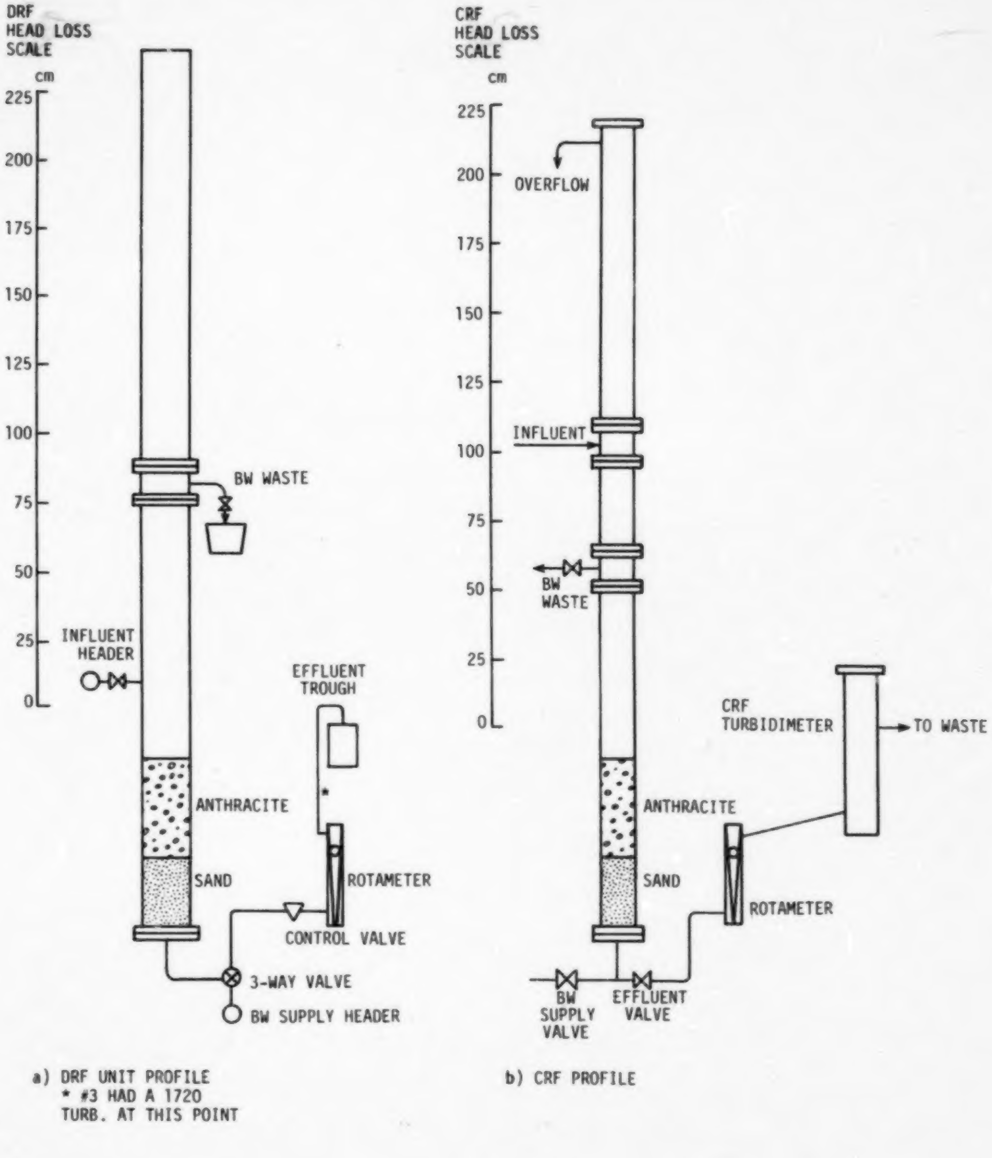

MAIN SPLITTER BOX
TO WASTE
ADJUSTABLE-WEIR-SPLITTER BOX
4-CELL FLOCCULATOR
TO WASTE
TO WASTE
OVERFLOW TO WASTE
OVERFLOW TO WASTE
FILTER
FILTER
STATIC MIXER
CHEMICAL INJECTION
SEWAGE INJECTION
TURBIDIMETER
TURBIDIMETER
TURBIDIMETER
TO WASTE
BACKWASH OUTLET
BACKWASH OUTLET
ANTHRACITE
ANTHRACITE
TO WASTE
FLOW METER
FLOW METERS
SAND
SAND
BACKWASH TANK
AIR SCOUR
3/4 IN. HOSE
RAW WATER PUMP
BACKWASH PUMP

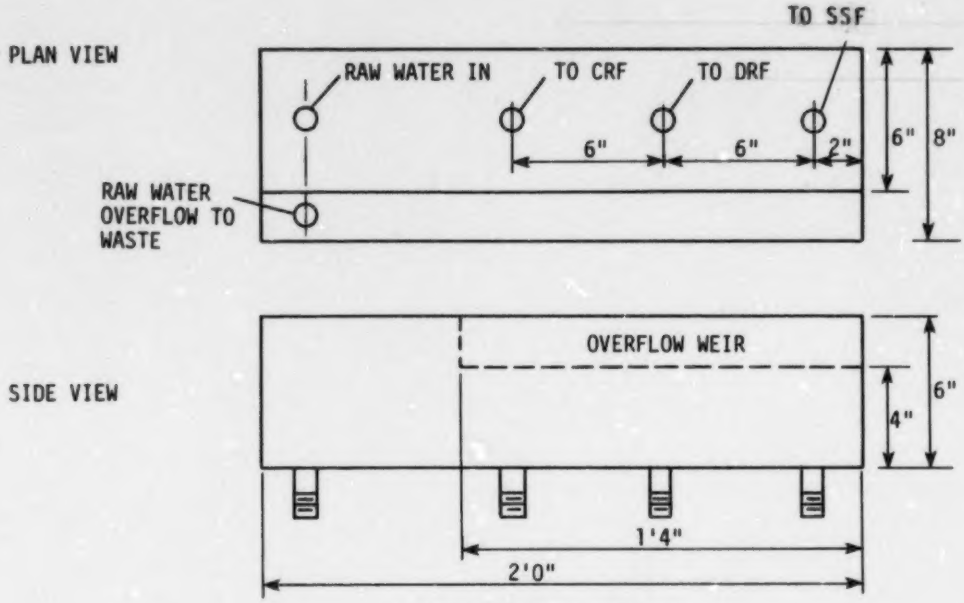
Fig. 3. Pilot plant schematic showing two constant-rate filters, one receiving flocculated water and one receiving unflocculated water in Phase III.


DRF
HEAD LOSS
SCALE
cm
225
200
175
150
125
100
75
50
25
0
INFLUENT
HEADER
BW WASTE
EFFLUENT
TROUGH
*
ANTHRACITE
SAND
ROTAMETER
CONTROL VALVE
3-WAY VALVE
BW SUPPLY HEADER
CRF
HEAD LOSS
SCALE
cm
225
200
175
150
125
100
75
50
25
0
OVERFLOW
INFLUENT
BW
WASTE
CRF
TURBIDIMETER
TO WASTE
ANTHRACITE
SAND
ROTAMETER
BW
SUPPLY
VALVE
EFFLUENT
VALVE

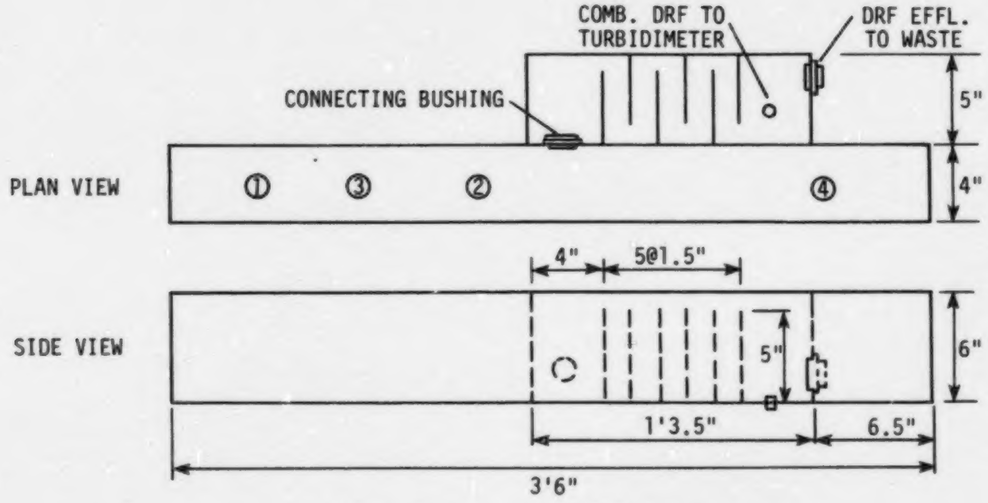
a) DRF UNIT PROFILE
* #3 HAD A 1720
TURB. AT THIS POINT

b) CRF PROFILE

Fig. 4. Filtration units (Head loss scales and filters are drawn to same vertical scale).


PLAN VIEW
RAW WATER IN
TO CRF
TO DRF
TO SSF
RAW WATER OVERFLOW TO WASTE
6"
6"
2"
6"
8"
SIDE VIEW
OVERFLOW WEIR
4"
6"
1'4"
2'0"

a) INFLUENT SPLITTER BOX


COMB. DRF TO TURBIDIMETER
DRF EFFL. TO WASTE
CONNECTING BUSHING
5"
PLAN VIEW
①
③
②
④
4"
4"
5@1.5"
SIDE VIEW
5"
6"
1'3.5"
6.5"
3'6"

b) DECLINING-RATE FILTER EFFLUENT TROUGH

②- DESIGNATES LOCATION OF FILTER DISCHARGES.

Fig. 5. Filtration appurtenances.

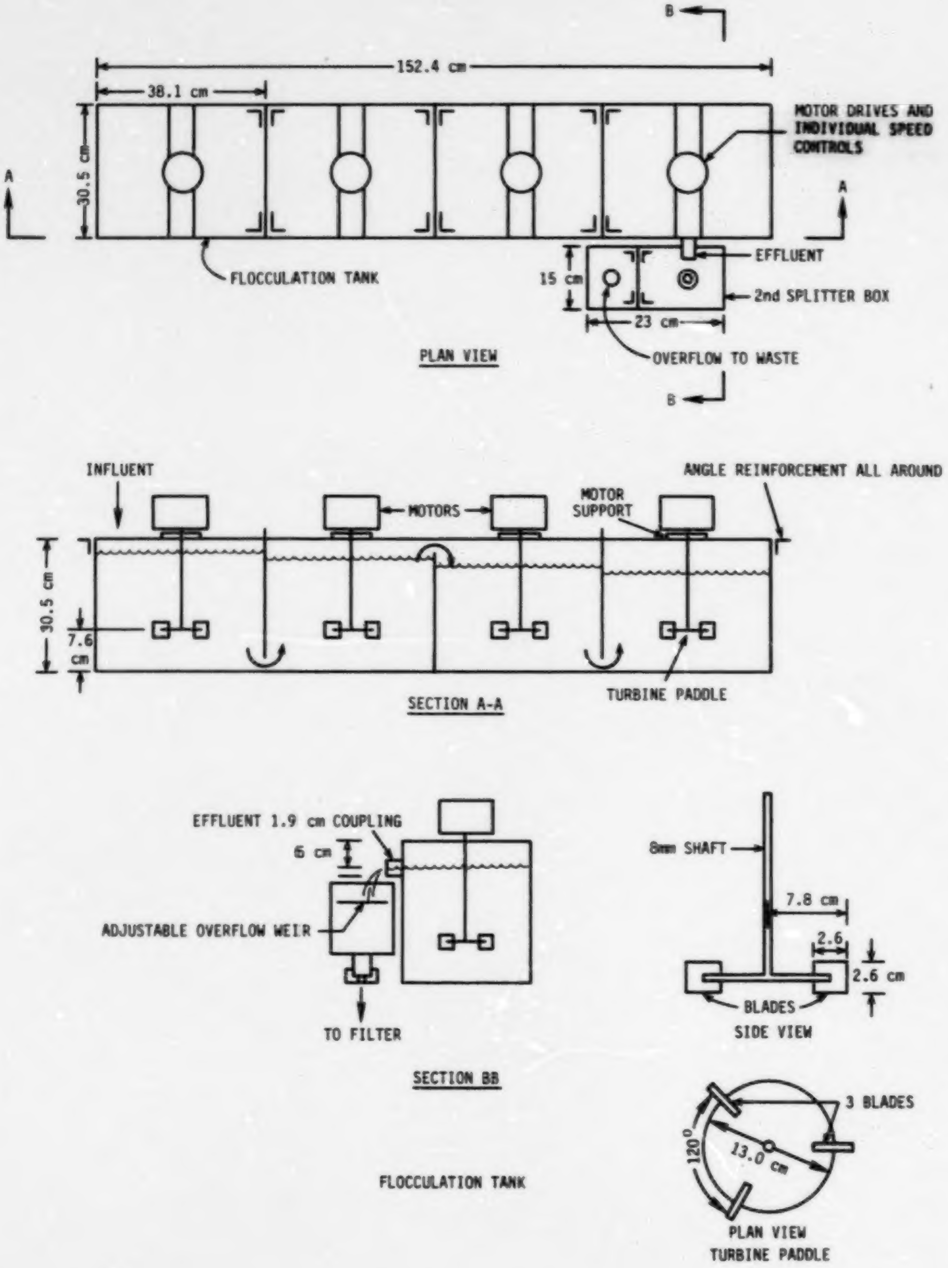
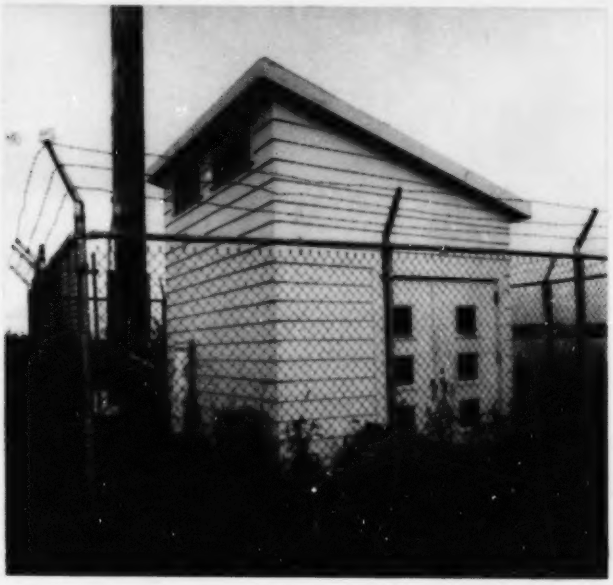
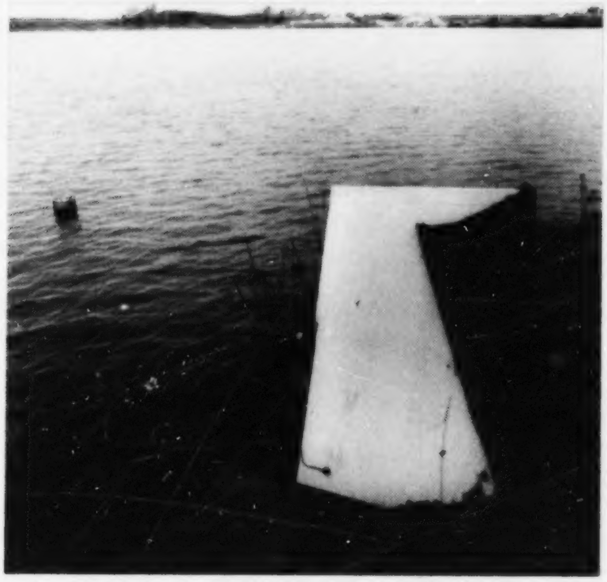

B
152.4 cm
38.1 cm
MOTOR DRIVES AND INDIVIDUAL SPEED CONTROLS
30.5 cm
A
A
FLOCCULATION TANK
15 cm
EFFLUENT
2nd SPLITTER BOX
23 cm
OVERFLOW TO WASTE
PLAN VIEW
B
INFLUENT
ANGLE REINFORCEMENT ALL AROUND
MOTORS
MOTOR SUPPORT
30.5 cm
7.6 cm
TURBINE PADDLE
SECTION A-A
EFFLUENT 1.9 cm COUPLING
6 cm
ADJUSTABLE OVERFLOW WEIR
TO FILTER
SECTION BB
8mm SHAFT
7.8 cm
2.6
2.6 cm
BLADES
SIDE VIEW
3 BLADES
120°
13.0 cm
PLAN VIEW
TURBINE PADDLE
FLOCCULATION TANK

Fig. 6. Flocculation tank and turbine paddles used in Phase III.

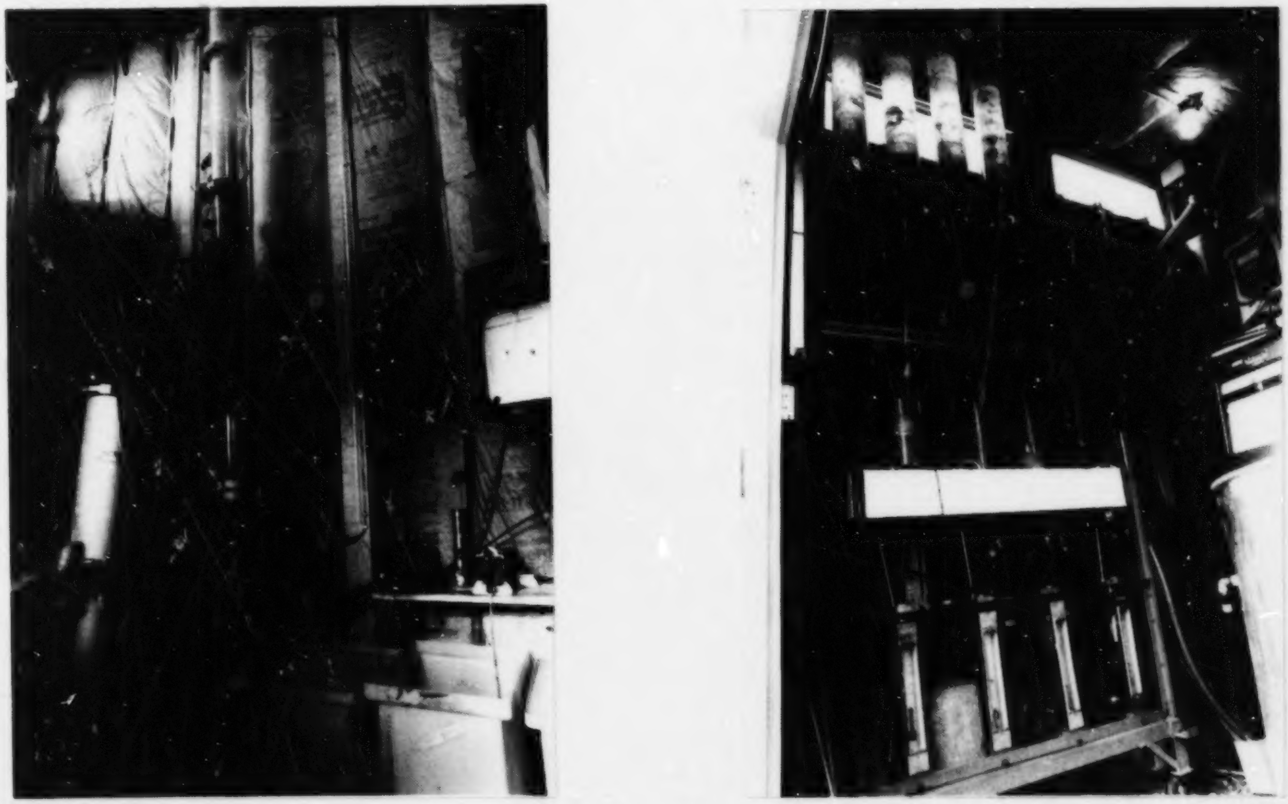


a) Pilot plant shelter.



b) Raw water intake (shelter covers the raw water pump).

Fig. 7. Raw water intake and pilot plant shelter.

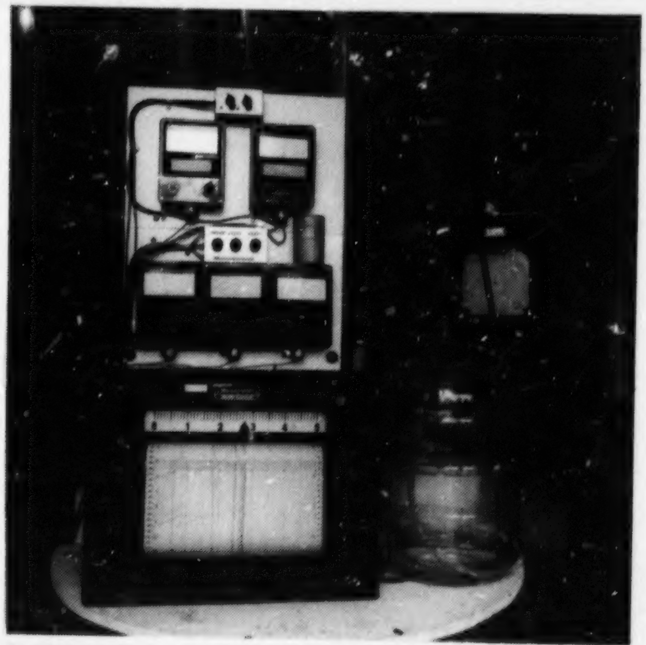


a) Constant-rate filter (1720 Turbidimeter is shown at bottom left, chemical feed equipment at bottom right).

b) Declining-rate filters (splitter-box is shown at top right).

Fig. 8. Filter equipment.

a) Chemical feed equipment with head loss recorder above (Ratio Turbidimeter, vacuum pump and pH meter are on the table).



b) Continuous recorder for turbidity (sitting on backwash tank).

Fig. 9. Miscellaneous equipment.

FILTERS AND PIPING

Slow Sand Filter

The slow sand filter shown in Fig. 1 consisted of a single cylindrical aluminum filter column with a 76.2 cm (30 in.) inside diameter and 2.74 m (108 in.) height. A splash plate was used in the filter above the sand surface to protect the media surface from disruption caused by incoming flow. The splash plate was 1.22 m (4 ft) above the bottom of the filter, thus about 3 cm above the initial sand surface prior to any scraping operations.

A single 1.27 cm (1/2 in.) diameter aluminum collector pipe was located with its outside about 0.6 cm (1/4 in.) off the bottom of the tank. It was capped at one end and threaded into a coupling at the effluent end of the pipe. The collector pipe had five orifices pointing downward. The 0.5 cm (3/16 in.) diameter orifices were drilled 12.7 cm (5 in.) apart on center.

To provide access for scraping the sand surface, an access opening with cover plate was provided. The opening was 46 cm high by 61 cm arc length (18 in. × 24 in.) with the lower edge of the opening 1.37 m (4.5 ft) above the bottom of the filter. The cover plate and flat rubber gasket were bolted to the housing. Raw water flow was delivered to the filter through a free-fall from the splitter box orifice to the prevailing water level in the filter. The water level was constantly rising as a filter run progressed.

Constant- and Declining-Rate Filters

The constant-rate filters were four-inch I.D. Plexiglas units with a screened filter bottom. Each unit was 288 cm, from the bottom of the media to the overflow outlet. The constant-rate filters had no effluent control devices and therefore were operated as an influent-flow-splitting filters.

The declining-rate filter bank consisted of four 15 cm (6 in.) I.D. Plexiglas units each with a screened filter bottom. Each filter was 328 cm from the bottom of the media (i.e., the screen) to the top of the unit. All four filters were bolted onto a structural steel frame. The filter bank included an influent header with 12 mm (1/2 in.) ball valves to each filter and an effluent header with three-way valves on each filter for alternate filter and backwash operations. Four rotameters were also bolted to the framework. See the Flow Measurement section for information on these rotameters.

The filter piping was as follows:

Constant-Rate Filters

Splitter-box to filter - 16 mm (5/8-in.) plastic hose
Filter to turbidimeter - 6 mm (1/4-in.) stainless steel tubing

Declining-Rate Filters

Splitter-box to filters - 51 mm (2-in.) Schedule 40 PVC
Influent header - 19 mm (3/4-in.) rubber hose

Effluent piping - 19 mm (3/4-in.) rubber hose and 9 mm (3/8-in.) stainless steel tubing

Waste and backwash piping for both CRF and DRF was either 16 mm (5/8-in.) plastic or 19 mm (3/4-in.) rubber hose.

Filter Media

The slow sand filter media and supporting gravel used in Phase I of this study are described in Table 3. The initial depth of sand was 0.94 m (37 in.).

TABLE 3. FILTER AND SUPPORT MEDIA USED IN SLOW SAND FILTER

Media Designation	Effective size d_{10} (mm)	Uniformity Coefficient	Depth (m)
Silica sand	0.32	1.44	0.94*
Silica sand	0.65 mm - 0.85 mm	--	0.048
Gravel (1/8 in. × 16 mesh)	--	--	0.053
Gravel (1/4 in. × 1/8 in.)	--	--	0.053
Gravel (1/2 in. × 1/4 in.)	--	--	0.051
Gravel (3/4 in. × 1/2 in.)	--	--	0.061

*Depth of silica sand varies due to scraping between filter runs.

The filter media used in the rapid filter during the various phases of the study are summarized in Table 4.

ANCILLARY EQUIPMENT

Pumps

During the first 8 months of Phase I the raw water pump was a 0.5 horsepower, 3450 rpm, 3.8 cm × 3.8 cm (1.5 in. × 1.5 in.) self-priming centrifugal pump. At the beginning of Phase II (with Phase I continuing in parallel), it

TABLE 4. FILTER MEDIA FOR RAPID FILTERS

Phase	Filter	Depth (m)	Effective Size (mm)	Uniformity Coefficient
I	One Rapid Filter (CRF)			
	Anthracite[*]	0.41	1.54	1.18
	Sand[†]	0.30	0.43	1.53
II	Rapid Filters (CRF & DRFs beginning 8/30/82)			
	Anthracite	0.35	1.40	1.36
	Sand	0.25	0.52	1.40
III	Two Rapid Filters (CRF)			
	Anthracite	0.46	1.40	1.36
	Sand	0.30	0.52	1.40

[*] Carbon Sales Inc., Wilkes-Barre, PA.

[†] Northern Gravel Co., Muscatine, IA.

became necessary to use a larger pump to deliver the larger flows to the four DRFs as well in the CRF and the slow sand filter. Therefore, a 1.5 horsepower, 3.8 cm × 3.8 cm (1.5 in × 1.5 in.) self-priming centrifugal pump was used. The backwash pump was a 0.5 horsepower, 3450 rpm, 3.8 cm × 3.8 cm (1.5 in. × 1.5 in.) self-priming centrifugal type. It was used to backwash the rapid-rate filters in all three phases.

Turbidimeters

Five continuous flow nephelometers were used for turbidity measurement at various times during the three phases of the study. They were used on the raw lake water, on up to two constant-rate filter effluents (one during Phases I and II, two during Phase III) and two on the DRF system in Phase II. One was located on the effluent of DRF No. 3 and the other on the blended effluent of the four DRFs.

All five of these nephelometers were Hach CR Low Range Turbidimeters, Model 1720, manufactured by Hach Chemical Company, Loveland, Colorado.

The turbidimeter used as primary standard for the research work was a Ratio Turbidimeter Model 18900 also manufactured by Hach Chemical Company. The five turbidity indicators attached to each of the continuous flow turbidimeters were Master Indicators also manufactured by Hach Chemical Company.

The turbidimeter output signal of the five turbidimeters was recorded on a six-channel continuous strip chart recorder. It was a 5 mv, full-scale Honeywell-Brown Electronik with chart speed of 5 cm (2 in.) per hour, manufactured by Brown Instrument Division of Minneapolis-Honeywell Company of Philadelphia, Pennsylvania.

Particle Counting Equipment

The particle counter assembly included a particle size analyzer, Model PC-320 with CMB-60 sensor, an automatic bottle sampler, and a digital printer, Model PO-500, all manufactured by the HIAC Division of Pacific Scientific Company of Claremont, California.

Head Loss Recorders

Continuous recording circular chart head loss recorders were Model 0202 manufactured by ITT Barton of Monterey Park, California. They had a range of 0-254 cm (0-100 in.) water column and recorded on a 25 cm (10 in.) diameter, 7-day, circular chart.

Flow Measurement

Raw water flow was measured with a rotameter, Series 700 Master Enclosed Flowrator (packing type, tube no. B6-35-10/77) by Fischer and Porter Company of Hatboro, Pennsylvania. The rated maximum flow was 36.4 L/min (9.6 gpm).

The slow sand filter and the constant-rate and declining-rate filter effluent flows were measured with Model 1112A rotameters manufactured by Brooks Instrument Division of Hatfield, Pennsylvania. All five rotameters had tube no. R-8M-25-5. The slow sand filter and constant-rate filters had float no. 8-RV-3 for a rated maximum flow of 2.95 L/min (0.78 gpm). The four declining-rate filter flow meters had two floats each. The smaller float was the same as that for the constant-rate filter. The larger float located below the smaller float was no. 8-RS-31 for a rated maximum of 9.80 L/min (2.59 gpm). During Phase III, the slow sand filter was no longer in service, so its flow meter was used on the second constant-rate filter placed in service during Phase III.

Vacuum Pump

The vacuum pump for filtering chlorophyll samples and for providing airwash pressure was a Millipore filter apparatus, Cat. No. XX6000000, manufactured by Millipore Filter Corporation of Bedford, Massachusetts. The filter paper for filtering chlorophyll samples was 4.25 cm glass microfibre filters (GF/C) from Whatman Ltd. of England.

pH Meter

The pH meter was a Corning Model 7 from Corning Scientific Instruments of Corning, New York.

Chemical Feed and Mixing Equipment

The sewage feed pump in Phases I and II was a 10-gallon-per-day Model CT-10 from Culligan USA of Northbrook, Illinois.

The alum or polymer feed pump during Phases I and II was a 22-gallon-per-day Everchlor Model AC-22 from Everpure Water Treatment Products of Westmont, Illinois.

Both of the above are variable-stroke-length, diaphram-feed pumps. The AC-22 pump had a stroke frequency of 22 strokes per min, and the CT-10 had a stroke frequency of 12 strokes per min.

During Phase III, the Model AC-22 was used for sewage feed, and a new feeder was used for alum or polymer feed. This feeder was a Series A Metering Pump, Model A 101-92 S, manufactured by Liquid Metronics Incorporated of Acton, MA. This feeder has adjustable stroke frequency and stroke length. It was generally operated at a fixed stroke frequency of 40 cycles per min and fixed stroke length. Dosage was controlled by changing the strength of solution being fed.

Mixing of the sewage and the chemical coagulants with the lake water was accomplished with two static mixers. In the fall and winter of 1981 during Phase I, the static mixer for the chemical feed was merely a small spiral coil of copper wire (about 6 mm diameter and 100 mm long) inserted in the influent hose to the filter as shown in Fig. 1. A commercial static mixer was used to mix the sewage with the flow. This static mixer was 19 mm (3/4 in.) diameter by 12.7 cm (5 in.) long PVC by Komax Motionless Mixer from Komax Systems, Inc. of Long Beach, California. The mixer contained 3 mixing elements.

Beginning in the spring of 1982 when the DRF system was being installed, the point of chemical injection was changed as shown in Fig. 2 so that the Komax mixer served both for chemical mixing and the sewage mixing. A separate discharge line from the pump to the splitter box was provided to supply the slow sand filter with water containing sewage but without chemicals. The splitter box was divided to serve the slow sand filter with the uncoagulated water.

The scale for weighing alum was a 2610 gram capacity Dial-O-Gram model from O'Haus. It was accurate to 0.1 gram.

Thermometers

The ambient air temperature and raw water thermometers were -50 to +50° C thermometers from Fischer Scientific Company of Pittsburgh, Pennsylvania.

Chemicals

Two types of alum were used on the project. The first type was technical ground aluminum sulfate (F.W. 666.42) from Fischer Scientific Company of

Fair Lawn, New Jersey. The second type was standard ground aluminum sulfate, meeting American Water Works Standard B403-70, from Allied Chemical Company of Morristown, New Jersey. The AWWA Standard B403-70 states that the alum shall contain "not less than 9.0% available water-soluble aluminum as Al" [5]. The change in types of alum was on August 12, 1982. Therefore, all of the Phase II data presented later were conducted with the standard ground alum. The polymer used in the project as a primary coagulant was Cat-Floc T with a molecular weight of about 50,000 from Calgon Company of Pittsburgh, Pennsylvania. In a few filter runs, alternate cationic polymers were tried including Cat-Floc T-1 and Culligan F-86, the latter marketed by Culligan International, Northbrook, IL.

The chlorine used intermittently as a disinfecting agent came from a variety of brands of bleach. The chlorine content of all bleaches was specified as 5.25% sodium hypochlorite.

The pH buffers used to calibrate the pH meter were as follows:

1. pH 4.00 +/- 0.01 at 25 C - 0.05 M potassium biphthalate, No. SO-B-101 from Fischer Scientific Company of Pittsburgh, Pennsylvania.

2. pH 7.00 +/- 0.01 at 25 C - 0.05 M potassium phosphate monobasic - sodium hydroxide, No. SO-B-107, also from Fischer Scientific Company.

3. pH 10.00 +/- 0.01 at 25 C - 0.05 M potassium carbonate - potassium borate - potassium hydroxide, No. SO-B-115, also from Fischer Scientific Company.

The sulfuric acid used for pH adjustment was standard 95-98% pure H_2SO_4, also from Fischer Scientific Company.

SECTION 7

ANALYTICAL PROCEDURES

PRIMARY PARAMETERS

Turbidity

Turbidity measurements of the raw water as it entered the shelter, and of the various filter effluents in service at any given time, were recorded continuously. During Phase II, the DRF No. 3 effluent stream was mixed with the other 3 DRF effluents in the effluent trough prior to sampling for the combined DRF turbidity reading.

During the filter runs, periodic indicator readings were recorded in the data book and noted on the recorder chart paper. Water samples were taken for turbidity measurements in the Hach Ratio Turbidimeter, which was used as the primary turbidity standard. In this way, a correlation could be made between the indicated turbidity and the actual measured turbidity. With knowledge of the prevailing correlation, additional turbidity values could be calculated from the turbidity charts generated by the recorder.

Particle Counting

Particle count samples were taken at critical times in the filter runs and counted on the Hiac Automatic Particle Size Analyzer. The analyzer had 12 channels that were set to specific particle size thresholds determined from the analyzer calibration. The threshold settings at various times during the research are presented in Table 5.

Each particle count sample was counted 3 times (3-60 mL aliquots). The automatic bottle sampler delivered 60 mL aliquots to the sensor. The sample time was maintained at about 8 min per 60 mL sample. The first count was discarded and the other two counts were averaged for the final values of particles per milliliter.

The CMB-60 analyzer sensor measured particles based on area of light blockage. As a particle passed between the light source and photo sensor, its light blockage area was converted to an equivalent particle diameter and recorded in the appropriate channel. Although actual particle sizes and shapes could not be determined by this method, it did indicate relative changes in particle size and number through the filtration process.

TABLE 5. THRESHOLD SETTINGS FOR AUTOMATIC PARTICLE SIZE ANALYZER FOR VARIOUS PROJECT PERIODS

Period	10/13/81-1/14/82		1/18/82-2/8/82		2/8/82-7/15/83	
Channel No.	Threshold Setting	Size µm	Threshold Setting	Size µm	Threshold Setting	Size µm
1	135A	<1	139A	1	139A	1
2	195A	2	197A	2	197A	2
3	285A	3	282A	3	282A	3
4	400A	4	398A	4	398A	4
5	550A	5.1	545A	5	545A	5
6	720A	6	719A	6	719A	6
7	900A	6.9	910A	7	910A	7
8	140B	8.9	142B	9	142B	9
9	265B	12.5	280B	13	240B	12
10	440B	16.5	455B	17	455B	17
11	600B	20	610B	20	610B	20
12	900B	25	912B	25	912B	25

The analyzer was operated in the "TOTAL" mode of operation, meaning that each of the 12 values given in the readout was the sum of particles of that threshold size and larger. Because of the interest in Giardia cyst-sized particles, the total number of particles between channels 7 and 9 (total in channel 7 - total in channel 9) was selected to indicate particles of that size range.

Flow

Flow measurement was made on the raw water flow and on the effluents from all the filters in service. Flow rates were observed on the raw water during Phases II and III in order to calculate chemical application rates and to ensure that the chemical dosage was constant during the filter run. Periodic effluent flow measurements were made during each run to note the flow, to ensure that desired flows were achieved, and to ensure in Phases II and III that equal area flows were delivered to the CRFs and DRFs being compared.

During Phase II, additional flow measurements were taken on each of the four DRFs during a series of backwashes to note the relative flow rates before, during, and after the filter backwash.

All of the flow measurements were readings on the 0 to 100 scales on the rotameters. Prior to use of the rotameters, each was calibrated in the flow range of interest. This procedure is detailed in the Quality Assurance section.

Head Loss

Filtration head loss was measured on slow sand filter, the CRFs, and on the DRF system. A piezometer tube was tapped into the filter wall of the slow sand filter and the CRFs and into the middle of the DRF influent header. These taps were above the filter media and below the low water level of the respective filters. The piezometers consisted of 6 mm (1/4 in.) plastic tubing from the tap to a board next to the filter. A metal tape was attached to the board to indicate the head loss. The zero reading of the tape was set at the static level of the filter outlet. These piezometer tubes indicated the water level in the filters. Also, from a tee fitting in the tubing, another 6 mm (1/4 in.) plastic tube was connected to each of the head loss indicating/recording instruments used for the rapid filters but not the slow sand filter.

Total Coliform Count and Standard Plate Count

The total coliform and standard plate count determinations were conducted either in the Food Technology Department (Phases I and II) or in the Analytical Services Laboratory (ASL) of the Engineering Research Institute. The analytical procedures for both counts were taken from Standard Methods [76] and are as follows:

a. Enumeration of coliform bacteria - Membrane Filter Technique, Section 909A. Under paragraph 5, Procedures, the alternative single-step direct technique (subparagraph d) was used instead of the enrichment technique (subparagraph c). At least two (and often 3) membrane plates were prepared using a different sample volume for each plate. Sample volumes of 0.1 and 1.0 mL were usually used for influent samples and 25, 50, and 100 mL for filtered water samples.

b. Confirmation of coliform bacteria - Standard Total Coliform MPN Test Section 908A. Some colonies that were presumed to be coliforms in the membrane filter enumeration were subjected to confirmation using procedures outlined in Section 908A. The confirmation involved inoculation into lauryl tryptose broth tubes, Presumptive Test 908A.1; transfer of positive tubes to brilliant green lactose bile broth (BGLBB), Confirmed Test 908A.2.c; streaking of positive BGLBB tubes on EMB agar, Completed Test 908A.3; transfer of typical or atypical colonies to lauryl tryptose broth, transfer of positive tubes to nutrient agar slants, and Gram staining of growth from the slants. Depending upon the number of samples being enumerated, from 5 to 10 colonies per week were confirmed by the above procedure.

c. Standard plate count - Section 907. Two plates of identical dilution were prepared for each Standard Plate Count.

Control plates were used for both total coliform count and standard plate count. Each new batch of dilution water was plated to ensure that there was no bacterial contamination. The dilution water source was checked for conductivity and copper concentration. Both parameters were determined to be within acceptable ranges.

SECONDARY PARAMETERS

Temperature

Two thermometers were in use on the project site. A thermometer attached to the north side of the shelter was used to make daily readings of ambient air temperature. A second thermometer was used to measure raw water temperature periodically at the point where the raw water entered the shelter. Measurements were recorded to the nearest 1° C.

pH

Measurements were made of the pH of the raw water and filter effluent. The readings were recorded to the nearest 0.01 pH unit. The saturated calomel electrode of the pH meter was kept full of saturated KCl solution.

At least once each day the pH meter was standardized. Depending on the expected pH of the sample, either the 7 and 10 buffers or the 7 and 4 buffers were used. First, the pH meter was standardized at pH 7.00 with the calibration knob and then standardized on either pH 4.00 or pH 10.00 with the temperature knob. About once per week, after standardizing the pH meter as outlined above, the pH of the 3rd buffer (either 4 or 10) was measured to check the performance of the machine. If the measured pH was off more than 0.2 pH units, corrective action was taken. In making the pH measurement, a 50 mL beaker was filled with the water sample and the pH electrode immersed in it. This water was then discarded and a second water sample taken for the actual pH measurement. The sample was swirled around the electrode momentarily and the pH reading was taken when the needle equilibrated.

INTERMITTENT PARAMETERS

Raw Water Quality

Every two weeks, the raw water was analyzed by the Analytical Services Laboratory (ASL) of the Engineering Research Institute (ERI). Analyses were done for alkalinity, total hardness, specific conductance, suspended solids, total PO_4, ammonia nitrogen, nitrate plus nitrite nitrogen, Kjeldahl nitrogen, soluble silica, chemical oxygen demand, and the chlorophyll series. Less frequently, a total dissolved solids (TDS) test was run to obtain data for a specific conductance vs TDS correlation.

The samples were taken in two plastic bottles. One bottle had about a 0.5 L volume and was fixed with 2 mL of sulfuric acid to preserve the sample for total and ortho phosphate tests. The second bottle had about a 4 L volume and was used for the remainder of the analytical tests. Both bottles were refrigerated until the analyses were performed.

Chlorophyll analysis was done on the raw water at the same time as the comprehensive raw water quality tests. In addition, raw water and filter effluent chlorophyll samples were collected during selected filter runs. Chlorophyll analysis was also done by the ASL. Each chlorophyll sample was prepared by passing a known volume of water through a glass fiber filter paper. The filter paper was placed in a dessicator in a freezer until the analysis. The analytical methods for raw water quality are found in Table 6.

TABLE 6. RAW WATER QUALITY ANALYSIS METHODS

Parameter	Analytical Method Reference
Alkalinity	Part 403 [77]
Total Hardness	Part 314b [77]
Specific Conductance	Part 205 [77]
Suspended Solids	Part 209d [77]
Total Dissolved Solids	Part 209c [77]
Total PO_4	Method 365.4 [59]
Ortho PO_4	Method 365.1 [59]
NH_3 Nitrogen	Method 350.1 [59]
NO_2 + NO_3 Nitrogen	Method 353.2 [59]
Kjeldahl Nitrogen	Method 351.2 [59]
Soluble Silica	Part 425c [77]
COD	Part 508a [77]
Chlorophyll	Part 1002G.1 & 1002G.3 [77]

Algae/Diatom Bioassay

Several times during the research, problems were encountered with algal blooms. The diatoms were specifically identified once (Nov. 1981) and the entire algae population was assayed twice in 1982. An insufficient sample volume was provided for the diatom analysis so the diatom numbers could not be determined. However, six genera of diatoms were identified in November 1981: *Synedra*, *Cyclotella*, *Navicula*, *Cymbella*, *Nitzschia*, and *Gyrosigma*. The first four listed have been cited as being of a filter clogging nature [77]. *Synedra* appeared to be present in the greatest quantity.

The two algae samples were collected in late July and mid-August 1982. For both samples, the algae enumeration was broken down by genera and the dominant taxa were identified by David Millie, a graduate student in the Botany Department of Iowa State University. For each analysis, a 2 to 3 L representative raw water sample and a concentrated plankton net sample were collected. The concentrated plankton net sample was used to identify genera and also species when possible. A 1000 mL aliquot of the raw water sample was taken and "fixed" to enumerate the algae taxa. The enumeration procedure came from Chapter 19 of *Handbook of Phycological Methods* [78] and is summarized below:

a. Place 1000 mL of sample into a graduated cylinder.

b. Add 10 mL of Lugol's iodine. The Lugol's iodine will preserve the algal suspension and help it to settle. Let it set for 3 to 4 days.

c. Decant liquid, leaving about 25 mL of volume.

d. Pipette a portion of the remaining concentrate into a Palmer-Maloney nannoplankton cell that has a volume of 0.1 mL.

e. Count the number of plankters.

The following equation was used to calculate the number of plankters per mL:

$$\text{Standing Crop} = \frac{(\#\ \text{organisms})(\text{cell area})}{(\text{area/field})(\#\ \text{fields counted})(0.1\ \text{mL})} \times \frac{(\text{vol of concentrate})}{(\text{original volume})}$$

This equation was developed by David Millie. It is a variation of the equation found on page 947 of Standard Methods [77].

SECTION 8

QUALITY CONTROL

OVERVIEW

The quality control procedures were developed to reduce or eliminate errors due to equipment, equipment handling, and analytical procedures. Bound data books were used to record all data including quality control information. The field documentation included date, time, chemical stock tank volume, sewage tank volume, raw and effluent turbidity readings on the Hach Model 1720 and Ratio Turbidimeters, flow measurement on the raw water and filter effluents, head loss, raw water and effluent pH, water and ambient air temperature, and remarks. Entries were made in ink at the time that the readings were made. Separate data books were kept in the lab for particle count and bacterial work results, including calibration records.

PRIMARY PARAMETERS

Turbidity

The instruments used for turbidity measurements were Hach Model 1720 Continuous Flow Turbidimeters, each with connections to a master indicator and the strip chart recorder, and a Hach Ratio Turbidimeter used as the primary turbidity instrument for day-to-day control. Turbidity measurement quality control included instrument calibration, cross checks between the instruments, and accuracy evaluations by the Environmental Protection Agency (EPA) as discussed below.

Hach 1720 Turbidimeters--

The raw water turbidimeter and the effluent turbidimeters were calibrated with Formazin solutions prior to being put into operation. The credibility of this calibration was questioned because of the awkward procedure. Also, the Hach Model 1720 Operation and Maintenance Manual termed this type of calibration a "difficult task."

As a second attempt, the raw water and effluent turbidimeters were calibrated using a frosted glass reflectance rod which came with the instrument. However, soon after beginning field work on the project, it became evident that periodic adjustments of the Hach 1720 turbidimeters would be necessary. Therefore, the Ratio Turbidimeter was used as the primary instrument to periodically adjust the Hach 1720 turbidimeters so that the 1720 readings agreed with the Ratio Turbidimeter readings. These readjustments were made during periods of stable turbidity and were recorded. Readings with the reflectance rod were taken after the initial standardization with the Ratio Turbidimeter. No adjustments were made as a result of the reflectance rod readings.

Two maintenance procedures were employed to ensure accurate turbidity readings. Two of the older turbidimeter tube interiors were repainted with flat black enamel paint to minimize scattered light interferences. Also, the turbidimeter tubes were periodically flushed out and cleaned with a brush. The photocells and lens system were cleaned with tissues when dirt was visible.

Ratio Turbidimeter--

The quality checks on the Ratio Turbidimeter included periodic calibration with a prepared Formazin Standard, more than once-weekly checks with a sealed Latex turbidity standard purchased from Hach Chemical Company, and three accuracy evaluations by the EPA. At the beginning of the research project in September 1981, the Ratio Turbidimeter was taken to Hach Chemical Company at Ames, Iowa, where a capacitor was added to dampen the erratic readout. At that time, the standardization and linear response were also checked. On September 25, 1981, October 12, 1981, March 10, 1982, May 24, 1982, October 29, 1982, March 24, 1983, and May 10, 1983, the Ratio Turbidimeter was calibrated using the procedure outlined in the Ratio Turbidimeter Operation and Maintenance Manual. A fresh 400 NTU Formazin solution was made up by the ASL prior to each calibration. The Formazin solution was prepared according to Section 214A of Standard Methods [77].

Throughout the weeks following the ratio calibration, the Latex standard was used to check the turbidimeter calibration. A fresh Latex standard with a shelf life of at least a year, according to Hach Chemical Company, was obtained early in the project. If the observed reading wandered 0.2 to 0.3 NTU from the Latex reading recorded at the previous calibration, a new Ratio Turbidimeter calibration was required.

During the pilot plant operation (Oct. 1981 to July 1983), the EPA conducted three accuracy evaluations. Each time, the EPA sent ampules of turbidity suspension. Without advance knowledge of the turbidity, each ampule was diluted according to instructions and a turbidity reading was made. The results were sent to the EPA and an accuracy rating was returned. The readings on all three sets of samples were accurate within the limits set by the EPA as shown in Table 7 below.

Particle Count

The equipment used for particle counting included the Hiac Particle Size Analyzer, automatic bottle sampler, and digital printer. The quality control associated with this equipment involved calibration of the particle size analyzer and controlled sample collection and dilution techniques. The particle size analyzer was calibrated on January 14 and August 10, 1982, with a sphere suspension of known particle size distribution obtained from the Hiac Company. The Hiac Company calibration solution was dated November 19, 1981

TABLE 7. TURBIDITY QUALITY CONTROL CHECKS USING EPA STANDARDS

EPA Sample Number	Date ISU Report	Sample Number	ISU Measured Turbidity (NTU)	EPA True Value (NTU)	Reported Acceptance Limits (NTU)
WS009	10/16/81	1	1.15	1.0	0.63-1.3
		2	6.70	6.7	6.1 -7.3
WS010	5/24/82	1	1.49	1.35	0.95-1.7
		2	5.50	5.5	4.6 -6.4
WS012	5/16/83	1	5.40	5.9	5.20-6.62
		2	0.34	0.42	0.18-0.66

and was to expire on November 19, 1982. The standard method for field calibration of the automatic particle size analyzer was followed as given in the operation and maintenance manual included with the analyzer [64]. The particle size analyzer channel settings (Table 5) were changed slightly as a result of the January 14, 1982 calibration but no change was necessary during the August 10, 1982 calibration. The channel 9 threshold was set at 13 μm on January 14, 1982 and reset at 12 μm on February 8, 1982 to better reflect the typical size of Giardia cysts.

Five hundred mL glass bottles were used as sample containers in the particle analysis. These bottles were prepared in the laboratory. They were washed with Alconox detergent (Alconox Inc., New York, N.Y.) and water, immersed in a sonic bath, and finally rinsed with distilled water. A clean synthetic film was placed between the bottle and the screw cap. Both nylon films* and Parafilm films† were used in this capacity. Prior to sampling, the bottles were rinsed with the water being sampled to remove the rinse water before drawing the final sample.

Bottles used for sampling in the field were also used for analysis in the lab. Transfer between containers was done only when dilutions were necessary to stay within concentration limits of the particle-counting sensor. The total particle concentration cannot exceed 12,000 particles per mL. When dilutions were necessary, all glassware was first washed with detergent and water, then rinsed with distilled water and a special deionized water which was also used for dilution water. The special deionized water was obtained by filtering building distilled water through Whatman No. 5 filter paper and a mixed bed deionizing column. In general, dilutions were only used for influent samples. Particle analysis was conducted on both the diluted sample and the dilution water. Through computations, the particle count of the original sample was determined from these data.

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*RCAS 2400 antistatic nylon films, RC cleanliness level, manufactured by Richmond Division of Dixico Incorporated, Redlands, California.

†Parafilm "M" laboratory film, manufactured by the American Can Company, Greenwich, Connecticut.

Particle count samples were refrigerated immediately after sampling to hinder particle alteration. Particle analysis was carried out as soon as possible. Counting delays ranged from 1 to 72 hours, although the majority of the samples were analyzed within 48 hours of sampling. An evaluation of the effect of delayed particle analysis was conducted in the following manner. Several replicate samples were drawn from two batch samples, A and B, which were collected on Sept. 29, 1982 in a pair of 2 L Erlenmeyer flasks. All samples were taken and refrigerated at the same time. At various time intervals one sample from both batch A and batch B were analyzed. The results of the time delay are shown in Table 8. Alteration of particle counts was evident despite preservation attempts by nonilluminated refrigeration. This variation is more extreme for the total count than for the count between channels 7 and 9. Since the particle count data between channels 7 and 9 were of primary interest and the counting was usually accomplished within 48 hours of collection, it appears that the particle count data reported was not adversely affected by the delay in counting.

TABLE 8. EVALUATION OF DELAYED PARTICLE ANALYSIS ON BATCH EFFLUENT WATER SAMPLES

		Particle Count	
Batch Sample	Delay Time (h)	Channels 7-9 (particles/mL)	Channels 1-12 (particles/mL)
A	7	51	1756
	29	67	1785
	54	49	1520
	124	38	1200
B	7	43	1710
	29	60	1709
	54	61	1312
	124	38	1260

The trends for sample A of Table 8 reflect the behavior of each of the individual channels. There was an increase in each channel after 29 h, except channel 1 and 2, and then a gradual decrease in each channel at 54 and 124 h.

The Automatic Particle Size Analyzer was switched on five minutes prior to counting to assure a low machine "noise." Sixty mL samples were drawn through the particle counter until three acceptable counts were obtained. The count was considered acceptable when the flow rate through the sensor and the total particle concentration of the samples were kept within the limits of 8 ± 0.8 mL/min, and below 12,000 particles/mL, respectively.

Flow

The six rotameters used for flow measurement were calibrated prior to use. Known volumes were timed through each rotameter for the range of flows of interest. Three replicate measurements were made at each flow rate. A calibration curve was made for each rotameter and subsequently used for all flow values read on that rotameter. The rotameter flow tubes and floats were cleaned periodically to ensure accurate results.

In Phase I, since the slow sand filter and single CRF were operating at different filtration rates, it was not necessary to ensure any equality between the filters. Rather, it was necessary only to keep the flow tubes and floats clean to obtain a dependable flow reading.

In contrast, in Phase II and Phase III it was vitally important that the filtration rates (i.e., flow/surface area) were identical so that valid comparisons could be made. In spite of the fact that the orifice caps used to split the flow were carefully machined with tapered holes to give a smooth sharp edge orifice flow, and were cleaned frequently using a test tube brush, some differences in filtration rate did occur and corrective action was required.

In Phase II (CRF vs DRF study), a 2.5 cm (one in.) ball valve was installed on the splitter box nipple serving the CRF and the drilled orifice cap was mounted below the ball valve. Thus, minor adjustments in orifice head could be made with the ball valve to bring the two filtration rates to equality.

In Phase III (CRF with and without flocculation), the flocculator was provided with a constant large flow by the main splitter box which gave a theoretical detention of 14 min in the flocculator. Only a small portion of this flow was needed for the CRF that followed. Therefore, a second small splitter box with adjustable head was fabricated and installed on the flocculator effluent. The majority of the flocculated water went over the overflow weir to waste. The desired amount for the CRF was split with an orifice cap operating under a constant head. If the two CRF flow rates were not equal, the orifice head was adjusted to bring the two filtration rates to equality.

Head Loss

The two recording head loss gauges were calibrated to actual piezometer readings. During Phase II, the piezometer calibration readings were from 0.3 to 1.7 cm higher than the recording head loss gauge readings on the constant-rate filter and from 0.2 cm lower to 1.7 cm higher than the recording head loss gauge readings on the declining-rate filter. During Phase III, the DRF recording gauge was used to record the head loss of the CRF without flocculation. The recording head loss gauge readings were periodically compared to the actual piezometer readings to ensure that there was no drift in the recording head loss gauge readings. The minor differences between the piezometer readings and the recording head loss gauge readings were not considered a problem. Sufficient piezometer readings were usually taken during filter runs so that the head loss data presented herein are largely from the piezometer readings. The recorded head loss chart data were used to fill in data during time intervals when no piezometer data were recorded, and to determine the time of the end of run when head loss reached the overflow level. The recorded charts were also useful to diagnose any problems which may have occurred when no one was present at the pilot plant, such as power failures, flow blockages, or low flow periods. These were all clearly evident on the charts and permitted rejection of runs where such events occurred.

Total Coliform Count and Standard Plate Count

As indicated in the Analytical Procedures section, the total coliform count and standard plate count determinations were carried out according to Standard Methods [77]. In addition, the following information was recorded:

a. Sample identification.

b. Date and time of sample collection.

c. Time interval between collection and analysis of sample.

d. Daily incubation temperature.

e. Temperature and pH of water sample when collected.

f. Source of media and lot number.

g. Observations of plates of uninoculated media to check sterility of materials for each set of samples. Observations on sterility of rinse water for coliform determination and dilution water were included.

h. Observations of known cultures of coliforms for typical reactions on media that were used for enumeration and confirmation.

i. Confirmation of a random selection of colonies having a sheen on the membrane filters by observation of typical reactions in lauryl tryptose broth (35° C for 48 hours) followed by incubation in BGLB broth for 48 hours at 35° C and by Gram staining. During Phase I and II, 5 to 10 typical colonies were selected each week for these confirmatory tests. A total of 412 colonies were chosen, of which 331 were confirmed as coliforms.

j. Analysis of the distilled water supply used for dilution blanks and rinse water for conductivity and copper content.

Chemical Feed

The quality control of chemical feed involved checking the chemical feeder flow and basing the stock solution requirement on that current feed rate. Also, since stock tank volume readings were recorded several times during a filter run, the actual volumetric feed rates for a portion of a run or the entire run could be calculated. Dry chemicals were measured using the field balance. When alum was used, at least 1/3 mL of sulfuric acid was added per L of stock tank volume to achieve a pH of about 3, thus ensuring that the alum was fully dissolved. In filter runs where pH was to be lowered intentionally to some desired level in the filtered water, greater quantities of concentrated sulfuric acid were added to the stock tank. Liquid polymer chemicals were measured with a 10 mL pipette graduated in 0.1 mL increments.

SECONDARY PARAMETERS

Temperature

Near the end of Phase I in November 1982, the two thermometers were tested by the Analytical Services Laboratory (ASL) for accuracy by establishing the ice point for each thermometer. The desired accuracy was +/- 1° C. The interior thermometer used for water temperatures was found to have an ice point of 0.0° C and was judged to be acceptable. The exterior thermometer used for ambient air temperature had an ice point of -2.9° C and thus the recorded air temperatures in the data books for 1981 and 1982 are presumed to be low by about 3° C. In 1983, a new thermometer was used for the air temperature which had an acceptable ice point.

pH

The quality control for pH measurements was obtained by the use of commercial buffers to bracket the expected pH and by frequent calibration of the pH meter against the buffers.

INTERMITTENT PARAMETERS

The quality control program of the Analytical Services Laboratory (ASL) is a routine practice in order to meet the requirements of all of their sponsoring agencies. The ASL has been approved by the Iowa Department of Environmental Quality under the provisions of the Safe Drinking Water Act. The ASL normally does three replicates for water quality parameters with the exception of most titrations (two replicates), conductance and chlorophyll, for which it is meaningless to make more than one instrument reading per sample.

SECTION 9

FILTRATION PROCEDURES

PROCEDURES DURING PHASE I

The flow scheme for the Phase I comparison of a slow sand filter and a rapid, dual-media CRF is best seen in Fig. 1.

Raw lake water was pumped to the filter shelter building. A portion of the raw water was passed through a continuous turbidimeter and wasted. The remainder of the water passed through an influent flow measuring device. After it was spiked with sewage, the remainder of the raw water and sewage passed through a static mixer; the mixture flowed into the splitter box. A uniform flow rate from the splitter box free-fell to two filters; thus, both filters were operated with influent flow splitting as a means of achieving constant-rate operation. Filter operating procedures were very simple using this operating system. For example, if a power failure or pump failure occurred, the water level would gradually fall toward the static level. However, the static level was above the filter media so that the media surface was not exposed. When flow was restored, the water level would gradually rise to the level existing prior to the interruption. There was no possibility for negative pressure within the media with this arrangement. Therefore, no problems were observed with gas binding of the filter media due to the release of dissolved gases.

The only operating problem observed with this system was an occasional reduction in inflow to the filter due to a fish scale or some other object partially blocking the flow splitting orifice. This became more of a problem in the winter of 1981-82. To correct this problem, a stainless steel, 10-mesh screen supported by a stainless steel angle iron frame was fabricated to cover the bottom of the flow splitter box. This screen would catch any large objects that could clog the flow splitting orifices. Periodic cleaning of the screen was necessary, especially when turbidity or algae levels were high in the lake water (about every two weeks under worst conditions).

Slow Sand Filter Operation

The slow sand filter was filled initially from the bottom to avoid entrapping air bubbles in the media. Filtered water from the backwash tank was used for this purpose. The filter was filled in this manner until the entire media depth was saturated. At this point, flow from the splitter box was directed to the slow sand filter, which started the first filter run.

During the filter run, raw lake water, filter influent water, and filter effluent water were sampled and analyzed for a number of different parameters as discussed previously.

The filter run ended when the water level reached the overflow of the filter near the top of the filter housing 149 cm above the original media surface (135 cm above the static level of the outlet). When the run ended the filter was drained until the water level was just below the media surface. The curved access cover was removed and the surface of the media was scraped. Scraping the filter involved removing the schmutzdecke, a gelatinous layer approximately one inch thick consisting of silica sand, microorganisms, and other particulates. After scraping, the access cover was refitted and a new filter run was started. After several runs were completed, the dirty sand that had been removed was washed and returned to the filter.

Filter media was washed by hand in a large bucket. The water and sand were agitated using a metal rod to dislodge the material that coated the sand grains. The dislodged impurities were washed away by the water stream produced by a garden hose.

Since the slow sand filter runs were fairly long, daily readings of flow, head loss, and raw and filtered water turbidity gave a good depiction of the filter performance. In addition, a continuous record of raw and filtered water turbidity was printed by the recording turbidimeter. Particle count and bacterial samples were taken of the raw lake water and the influent to the slow sand filter on at least a weekly basis. Effluent samples were taken at least biweekly, with a number of extra samples taken during the initial improvement period of the filter run.

Rapid Constant-Rate Filter Operation

Operation of the CRF involved several procedures, including the placement of the filter media, coagulant and dosage selection, filter operation and sampling, and backwashing.

Placement of Filter Media--

The placement of the dual media in the CRF involved the following steps. First the filter sand was placed to a depth about 2 to 3 cm (0.8 to 1.2 in.) deeper than the desired finished sand depth. The sand was backwashed and the backwash valve was shut off very slowly to allow the maximum degree of stratification of the sand, the finer grains accumulating at the surface. The excess sand was then siphoned off to yield the desired finished depth of sand, thus removing the finest sand grains. The anthracite was then placed in the same manner, backwashed, and skimmed by siphon to yield the desired total bed depth.

Coagulant and Dosage Selection--

It was decided at the onset of the project that only a single coagulant should be used because of the project emphasis on simple treatment methods for small systems. It was recognized that some direct filtration systems use combinations of metallic coagulants and polymers as primary coagulants, and may also use nonionic polymers as filter aids. However, for small systems with unskilled operation, the use of more than one chemical was considered unacceptable. Therefore, either alum or a cationic polymer was selected as the sole chemical coagulant.

In some alum coagulated runs, the pH was lowered to about 6.8 because of a presumption that this would be a favorable pH. For low alkalinity waters such as mountain waters, a pH of 6.8 would be achieved easily, and might be reached by the alum addition alone. However, for the Hallett's quarry water with alkalinity of about 150 to 250 mg/L as $CaCO_3$, achieving a pH of 6.8 required the addition of substantial amounts of concentrated sulfuric acid to the alum feed tank.

This no doubt would be an impractical procedure for a high alkalinity water, but it was done in this research because of the project emphasis on mountain waters and the desire to operate at a pH more typical of alum treatment of such waters.

The experimental procedure for the rapid CRF runs generally consisted of making one run per week that included the monitoring of turbidity, particle count, and bacterial count; these were called observation runs and usually began on Monday mornings. The observation runs were usually preceded by a few days of operation to select the apparent optimum chemical feed rate using turbidity monitoring alone to guide in that selection. This period usually occurred over the weekend or during the 3 or 4 days preceding an observation run. By Monday morning, the optimum dose had usually been selected and a formal observation run would commence. As much as possible, dosage changes were not made during an observation run, even if the lake water quality changed due to weather changes.

The period of preliminary operation prior to an observation run also allowed a few days to acclimate the filter media to the chemical treatment when a change in the type of chemical coagulant occurred.

The selection of alum dosage was based primarily on experience in trial runs preceding an observation run. It became evident that while an excessive alum dosage did not cause poorer filtrate quality in the early part of the filter run, it did cause more rapid head loss and earlier terminal breakthrough of turbidity. As alum dosage levels were increased for a given raw water situation, each increase gave some improvement in filtrate quality (after the initial improvement period) but the response was one of diminishing returns, i.e., less benefit per unit increase in alum feed. So the choice of the best alum feed was largely a compromise between the benefits of better filtrate quality early in the filter run vs the detriments of earlier breakthrough late in the run and higher head loss increase per unit time. Optimum alum feed was rather independent of raw water turbidity. For example, in many runs with 7 mg/L alum and pH about 7.8, the filtrate would remain nearly the same over raw water turbidity fluctuations from 2 to 20 NTU.

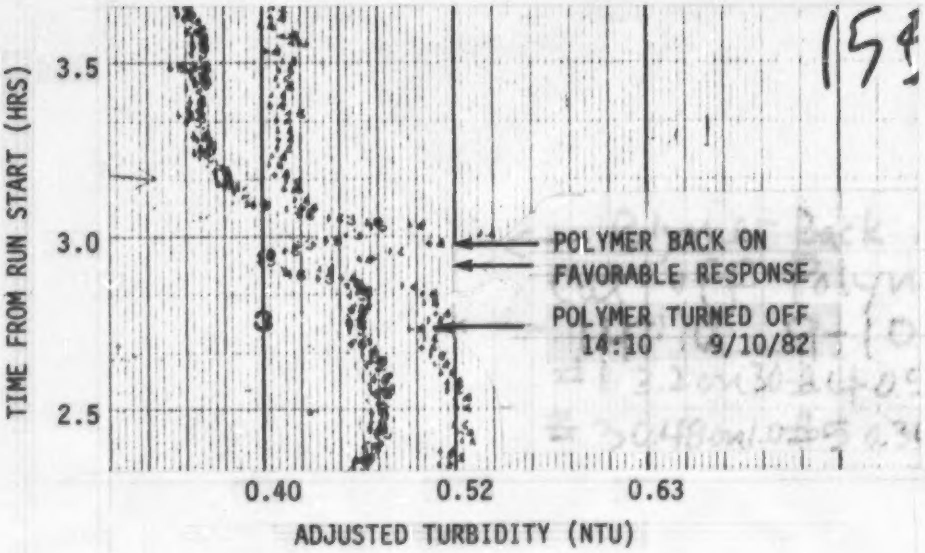
In contrast, the selection of dosage of cationic polymer as primary coagulant was more difficult because overdosing resulted in poorer filtrate quality. Therefore, the selection of dosage was based upon experience, and upon the response of the filtrate quality when a change of dosage was made. This was done by waiting until a steady filtrate quality was obtained (i.e., after the initial improvement period of a filter run). The dosage was then raised or lowered and the filtrate response was observed on the continuous turbidimeter recorder chart. If the change was favorable, after waiting for steady performance, a second change in the same direction was made, and so forth. If the response was unfavorable, a change in the reverse direction was made. This process was repeated until the apparent optimum dosage had been determined.

Late in the experimental work of Phases I and II, it was observed that the process of dosage selection for cationic polymers could be sped up by briefly stopping the chemical feed entirely. If the dosage had been too high (i.e., overdosed), the turbidity of the filtrate would immediately improve for a brief period of 5 to 10 min and then begin to degrade. It is presumed that the brief improvement is the result of reserve polymer in the filter and on the media surfaces. When the reserve is exhausted, the filtrate deteriorates. The procedure for evaluating the optimum polymer dosage is illustrated in Fig. 10. Near the beginning of Run J-4 the polymer feed was shut off, the effluent turbidities went down temporarily and then started to increase, as shown in Fig. 10a. This effluent turbidity reaction was interpreted as an overdose and possibly a slight polymer buildup in the media. Consequently, the polymer dosage was reduced from 0.53 to 0.35 mg/L. After the polymer feed was restarted at the lower dosage, the effluent turbidities stabilized at lower values.

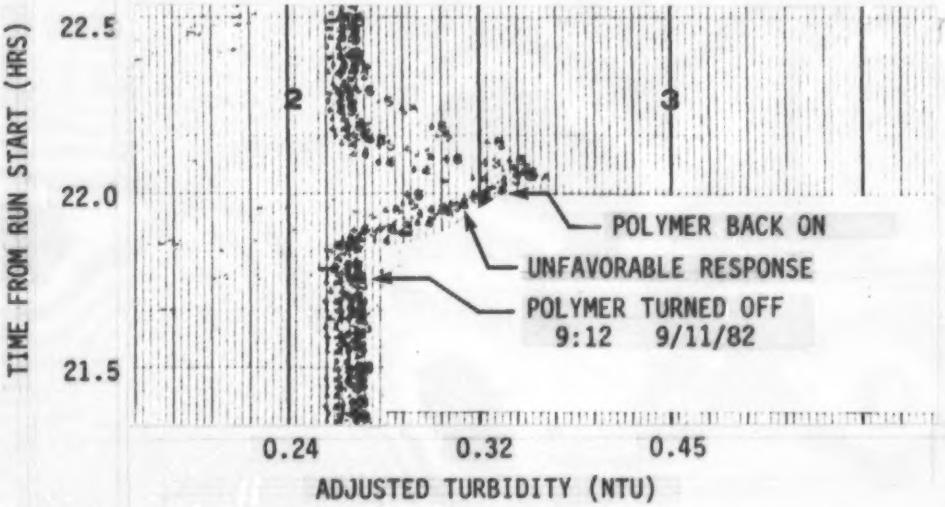
Figure 10b illustrates the effluent turbidity reaction when the polymer dosage was at or below optimum. When the polymer feed was shut off, the effluent turbidity increased. When the feed was restarted the effluent turbidity recovered. No change in polymer dosage was made in this case, although an increase in dosage could have been tried. This method of evaluating polymer dosage was possible because of the continuous turbidity record.

Filter Operation and Sampling--

The operation of the CRFs in Phases I and II using the influent splitting arrangement was simple and trouble-free. After a backwash operation was completed the filter water level was at the level of the backwash outlet. Since the influent flow was not shut off during the backwash operation, it continued to flow into the filter during and after the backwash. The effluent valve was opened fully and filtration began. If the equilibrium head loss for the clean filter at the prevailing inflow rate was lower than the backwash outlet level, the water level would fall to that equilibrium level in a few minutes as the run commenced. Then, as the filter got dirty, the level would gradually rise, ultimately to the overflow level which represented the maximum terminal head loss available. Some runs were terminated before reaching terminal head loss if turbidity breakthrough was severe.



a) Favorable response in effluent turbidity; polymer dosage was reduced.



b) Unfavorable response in effluent turbidity; no change, or increased dosage is indicated.

Fig. 10. Effluent turbidity responses to polymer coagulant shut off during Run J-4.

The only mechanical problems in the operation of the CRF were related to water flow rate or chemical or sewage feed rate. The flow rate problem was due to periodic blockage of the inlet splitting orifices, which was discussed earlier. The chemical or sewage feed were sometimes interrupted by an air bubble in the diaphragm of the pump or by a sticking check valve (poppet valve). The improper feed was detected by the level readings taken on the source tank. Corrective action was taken and the run was restarted if the coagulant feed rate had been off the desired dosage.

During all the filter runs, a continuous record of turbidity was obtained. The recording head loss instruments were installed on May 25, 1982, beginning with run E-1a; a continuous record of head loss was obtained for all subsequent filter runs. Prior to that date, head loss was read and recorded manually when any one was present at the pilot plant. This manual reading of the piezometers and manual recording was continued through the entire project, even after the recording head loss equipment was in operation.

During the observation filter runs, periodic samples were taken of the influent and effluent for particle count and bacterial analysis. The intended sampling plan was to take about 10 samples during the filter run, 3 or 4 during the initial improvement period, 3 or 4 during the period of optimum operation, and 2 or 3 during the terminal breakthrough period (if it occurred). This plan was not always fulfilled completely. In some cases, the runs would be shorter than anticipated and fewer than 10 samples would be collected. By the time the operator arrived at the pilot plant in the morning, the run would be over and the late run samples would have been missed.

Backwashing Procedure--

The backwashing procedure for the rapid filters was essentially the same in all three phases of the study. The objectives of the filter backwash were to clean the media and to prevent media problems such as mudballs and surface blinding which would add undesirable variables to the research. The backwash procedure is outlined below:

1. The filter effluent valve was closed.
2. The filter influent valve was closed (DRFs only in Phase II).
3. The head space above the filter was drained to below the backwash waste outlet through the backwash waste line.
4. The air compressor was turned on and the backwash valve was opened.
5. Air alone was delivered to the filter to provide a turbulent agitation of the media for about 3 minutes.
6. The air wash was shut off.

7. The backwash pump was turned on and the backwash valve was opened slowly to allow the media to fluidize.

8. The filter was backwashed for 7 to 8 min with 30 to 50% expansion.

9. The backwash flow was shut off slowly to allow the dual media to separate and settle. No special compaction procedures were practiced in the first 10 months of Phase I. However, they were necessary in Phases II and III and will be discussed later.

10. The effluent and influent valves were opened, making the filter operational again.

SPECIAL PROCEDURES DURING PHASE II

Replacement of Filter Media

At the beginning of Phase II, the filter media was removed from the CRF and new dual-media was installed in the CRF and the four DRFs. This was done because of the desire to place the media in the five filters in a controlled manner to try to avoid any differences in media between the filters. To do this, the sand was placed first, placing one scoop at a time into each of the five filters in succession (about 2 to 3 cm depth per scoop), then repeating this procedure until the desired total depth of sand had been placed. After backwashing and checking the depth, the anthracite was placed in the same controlled manner until an excess depth of 2-3 cm (0.8-1.2 in.) had been achieved. The filters were then backwashed and skimmed by siphoning off the surface coal to achieve the desired finished depth as discussed under Phase I procedures. In this case, inadvertently, the sand had not been skimmed and some fine sand grains were observed in subsequent filter runs to migrate all the way to the top of the anthracite.

A number of problems ensued in the early weeks of Phase II, especially in obtaining equality of head loss development between the CRF and the DRF system under identical operating conditions. Because of this, new filter media were again installed in all five filters on Aug. 28, 1982, as one step toward solving the problems. The procedure was identical to that described in Phase I procedures, including the skimming of both the sand and the coal to remove any undesired extra fine grains of media.

Overview of Phase II Operation

During the comparison of CRF vs DRF, the filtration runs were generally divided into three parts: the preliminary run, the formal run, and the secondary run. The preliminary run involved bringing the filtration apparatus into a steady-state condition wherein the CRF and DRFs could be compared. The formal run then was made with the collection of a full set of data. Subsequently, the secondary run was made. No particle count or bacterial samples were taken in the secondary run.

Preliminary Run

The purpose of the preliminary run was to bring the filtration plant into equilibrium. The chemical feed rate was optimized during this period using the procedures described earlier. This was difficult to do when the raw water quality was extremely variable. The areal flow rate to the CRF was matched with the areal flow rate to the DRF bank during this period. Adjustments were made in the position of the ball valve located above the CRF orifice. The initial head loss increases on the backwashed CRF and DRF bank were observed to ensure that there were no media compaction effects that would affect a head loss comparison. Runs J-5b and J-8a were strictly head loss comparison runs in which all filters were backwashed, recompacted by a uniform procedure and then started up with no filter backwashing during the comparison. Over a 14.5-hour-period in Run J-5b, the difference in head loss increase was 2 cm (52 cm on the CRF and 54 cm on the DRF bank). After 17 hours in Run J-8a, both the CRF and DRF developed 16.5 cm of head loss increase.

Finally, the preliminary run was used to bring the declining-rate filter bank into steady state operation. Starting with four clean DRFs, operation during the preliminary run continued until all four filters had been washed at regular intervals and a reasonably steady head loss pattern had developed. Usually after the fourth backwash, the head loss would increase between backwashes but would decrease an equivalent amount after the backwash. This steady state operation gave a fairly uniform jagged-tooth head loss profile typical of declining-rate filter operation.

The individual DRF flow was adjusted via the effluent control valves. A maximum flow of 1.5 times the average individual filter flow was desired for a clean filter when it was at level 3 [18]. Level 3 corresponds to the water level just before backwash. Level 4 is the maximum level which occurs in the other operating filters when one filter is being backwashed. Level 4 and Level 3 were estimated at 215 cm and 165 cm, respectively. For a target mean flow rate of all filters of 7.3 m/h (3.0 gpm/ft^2), the water level of each of the 4 DRFs while clean was maintained at 165 cm while the control valve was adjusted to provide a flow of 11.2 m/h (1.5 times the average individual filter flow). The control valve positions for the average 7.3 m/h flow rate were: 1) 3-1/8 turns from full open, 2) 2-3/4 turns from full open, 3) 2-13/16 turns from full open, and 4) 3-1/16 turns from full open. For the runs at a mean flow rate of 13.3 m/h, the control valves were kept in the full open position because even at 165 cm (Level 3), the filtration rate for the clean filters did not exceed 1.5 × 13.3 m/h.

Formal Run

At the start of the formal run, the CRF and the dirtiest DRF were backwashed, and both filtration systems were put into operation simultaneously. A full set of readings and samples was taken during the formal run. This included raw water and effluent chlorophyll samples during the middle portion of the run. During the initial improvement period on the CRF and DRF number 3, additional turbidity readings and particle count and bacterial samples were taken. The formal run was complete after 4 backwashes on the declining-rate filter. These washes were timed to yield an approximately steady state head loss pattern and so that the 4 filters would be washed during a single run of the CRF.

Secondary Run

At the completion of the formal run, when a DRF was being backwashed, the CRF was also backwashed. The CRF was again started simultaneously with the DRF bank and a second filtration comparison run was conducted. The procedure for the secondary run was the same as the formal run except that no particle count or bacterial samples were taken. Head loss and turbidity were the basis for comparison of the secondary runs. For the last 2 runs of Phase II, the secondary run came before the formal run.

Backwashing

The backwashing procedure in Phase II was identical with that described for Phase I with one exception. After some weeks of operation in Phase II, it was determined that the rate of head loss development was affected substantially by the manner of compacting the filters after the backwashing was completed. Therefore, a strict compaction procedure was found necessary to ensure equal head loss behavior between the two systems. The backwash valve was opened and shut to provide short bursts of water which lifted the media. Each time when the burst was over, the bed would compact slightly. This was repeated about 5 or 6 times until the bed had achieved nearly maximum compaction at which time a mark was made on the filter housing to guide future backwashes. Thereafter, the bed was compacted in the same fashion and to achieve the same bed depth after each backwash. After this procedure was instituted, similar head loss increases could be observed on the CRF and DRF bank when both were operated as constant-rate filters under identical conditions. The backwash rates for the CRF and DRF are presented in Table 9. The backwash rates came from actual flow measurements; simultaneous notations of the media expansion were recorded.

SPECIAL PROCEDURES IN PHASE III

The comparison of direct, in-line filtration with direct filtration with flocculation in Phase III generally used procedures identical to Phases I and II. Two CRFs were operated in parallel with identical filter media. New filter media were installed at the beginning of Phase III, using the same precautions to ensure equality of media for the two filters as described under Phase II Procedures. Chemical dosage, filter operation, and backwashing procedures were the same as in the prior phases. Bed compaction procedures were the same as described in Phase II Procedures to ensure that head loss comparisons would be valid.

The relationship between the rotating speed and torque for the turbine flocculator paddles was measured using a Cole-Parmer Model 4425 Master Servodyne manufactured by Cole Parmer Instrument Company, Chicago, Ill. Initially, each turbine paddle had six blades. However, even at the lowest possible operating speed of the drive motors, overheating of the motors would

TABLE 9. BACKWASH RATES AND BED EXPANSION IN PHASE II

Filter No.	Consolidated Media Depth (cm)	Media Exp. (%)	Backwash Rate (m/h)
CRF	59.7	31.0	75.0
		53.2	103.1
DRF #1	60.3	43.2*	84.5
DRF #4	60.1	44.7*	91.9
		30.0	75.7

*Maximum media expansion possible with the backwash supply pump used.

occur and the motors would stop intermittently. Therefore, after a few days of unsuccessful operation, the number of blades was reduced to 3 per paddle as shown in Fig. 6. The paddles were operated at 60 rpm for all 4 cells of the flocculation tank and in all subsequent filter runs. The torque measured at this speed corresponded to a velocity gradient (G) of 56 s^{-1} at 18° C. The flow rate through the flocculation tank was constant in all filter runs, providing a total theoretical detention time of 14 min. Thus, the dimensionless flocculation parameter $G \cdot t$ that resulted was 47,000.

In the initial filter runs of Phase III, some difficulty was experienced in achieving identical areal filtration rates for the two CRFs in service. To solve this problem, a second splitter box was fabricated and installed on the flocculator effluent as shown in Fig. 6. This splitter box had an adjustable weir controlled by a threaded bolt and wing nut. Small adjustments of the weir were made to bring the areal filtration rates to equality.

SECTION 10

RESULTS

RAW WATER ANALYSES

The raw quarry water was sampled and analyzed at approximately two-week intervals during the entire project. The raw water characteristics during this period are given in Table 10. TDS was measured during the project to establish a correlation with specific conductance. From 11 sets of TDS and specific conductance values, the least squares line was: Specific Conductance (micromhos/cm) = 1.52(TDS (mg/L)) - 14.06, with a coefficient of correlation (r) of 0.738.

A correlation was also found between suspended solids and turbidity. Nineteen sets of values were available for the period of November 9, 1981 to May 31, 1983. The least squares line was: SS(mg/L) = 1.83 × (Turbidity (NTU)) - 1.11 with a coefficient of correlation (r) of 0.978. All of the 19 turbidity values were less than 17 NTU. Only two turbidity values were greater than 5. Exclusion of these two data pairs resulted in the least squares line: SS(mg/L) = 1.25 × (Turbidity (NTU)) + 0.46 with r = 0.845. In subsequent calculations the latter equation will be used for turbidity less than 5 NTU and the former for turbidity greater than 5 NTU.

Raw water quality data were collected on Hallett's Quarry southeast cell for over a year prior to the start of this project in the fall of 1981. These data were analyzed by the same methods shown in Table 6 by the ERI-ASL at Iowa State University. However, pH and temperature measurements were made in the field. Figures 11, 12, and 13 show the seasonal averages and ranges for selected parameters. Figures 11 and 13 show the effect of algal blooms on pH and alkalinity. Chlorophyll-a is an indication of the total viable and non-viable algae concentration. Individual raw water chlorophyll-a measurements are also reported in Table 11. During the summer of 1982 when the chlorophyll-a level increased dramatically, the carbon dioxide and alkalinity levels decreased and the pH increased due to uptake of carbon dioxide by the algae. Figure 13 does not give the pH values prior to this project period because the pH was measured in the laboratory instead of in the field. The lag time between sample collection and measurement may have created an error. Figure 12 indicates the nitrogen form variations. The most obvious variations were during the summer of 1980 and the spring and summer of 1982. It is suspected that this was due to the increased rainfall and runoff into Hallett's Quarry during those periods. Note that there was no significant algae activity during the summer of 1981 as indicated by the low chlorophyll-a levels on Fig. 13.

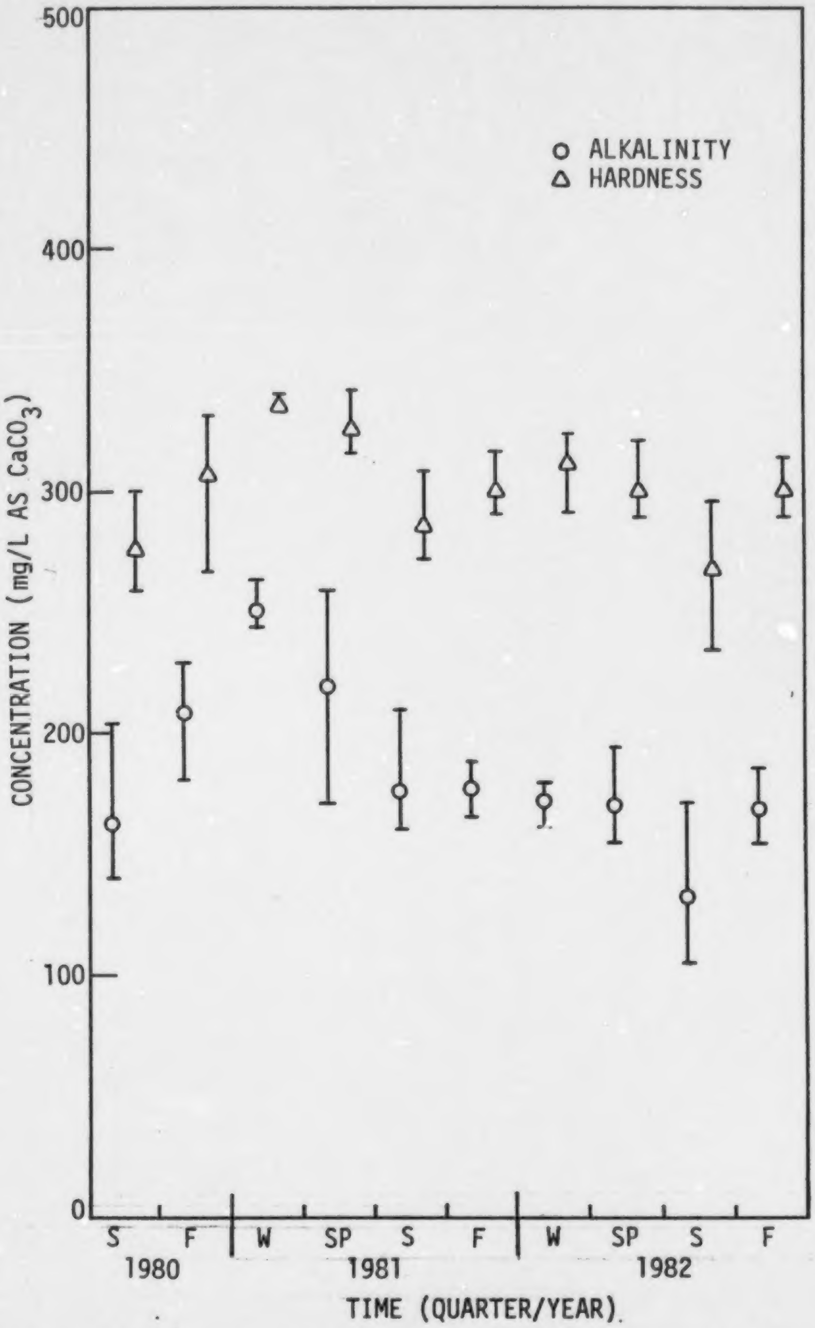


Fig. 11. Alkalinity and hardness in Hallett's Quarry southeast cell. Data point is mean value for the season and bar spans the range.

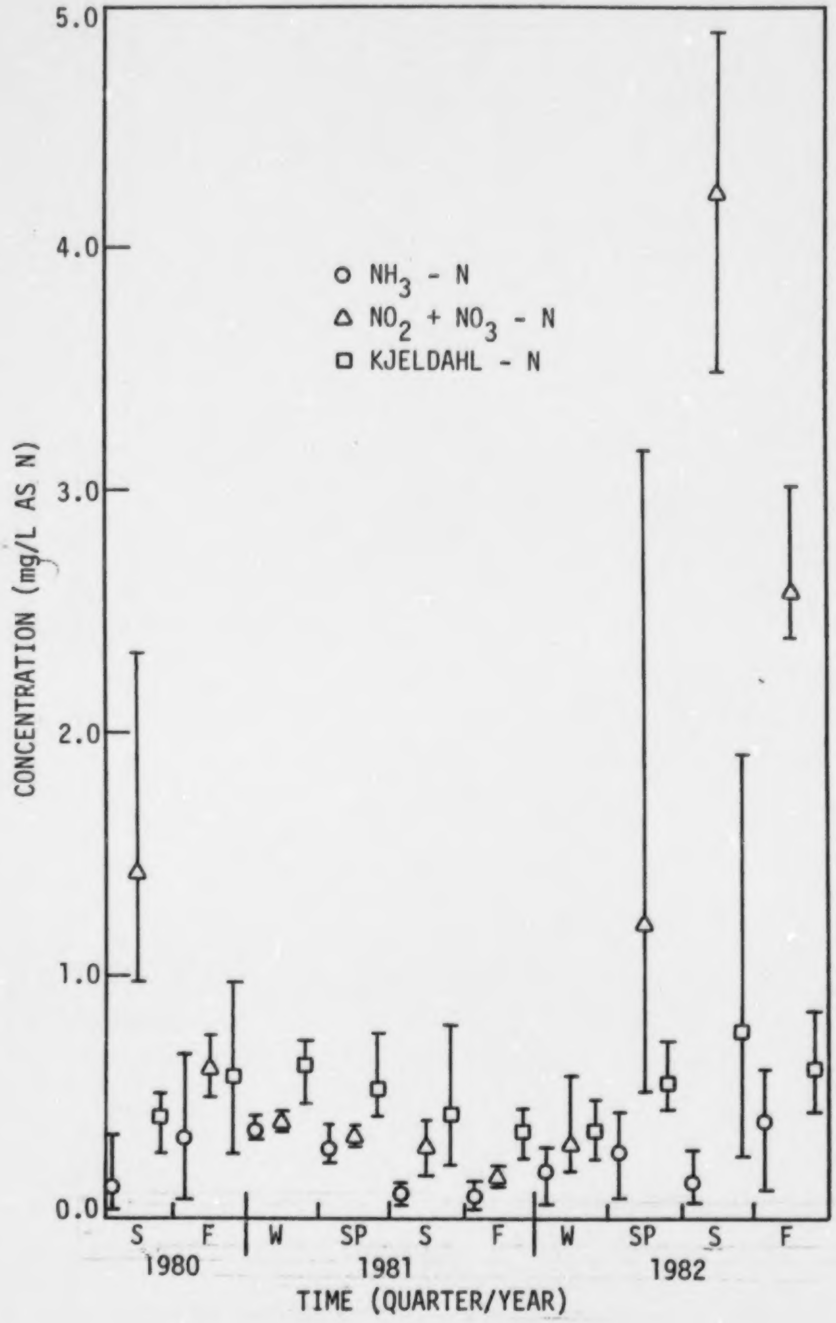


Fig. 12. Nitrogen forms in Hallett's Quarry southeast cell. Data point is mean value for the season and bar spans the range.

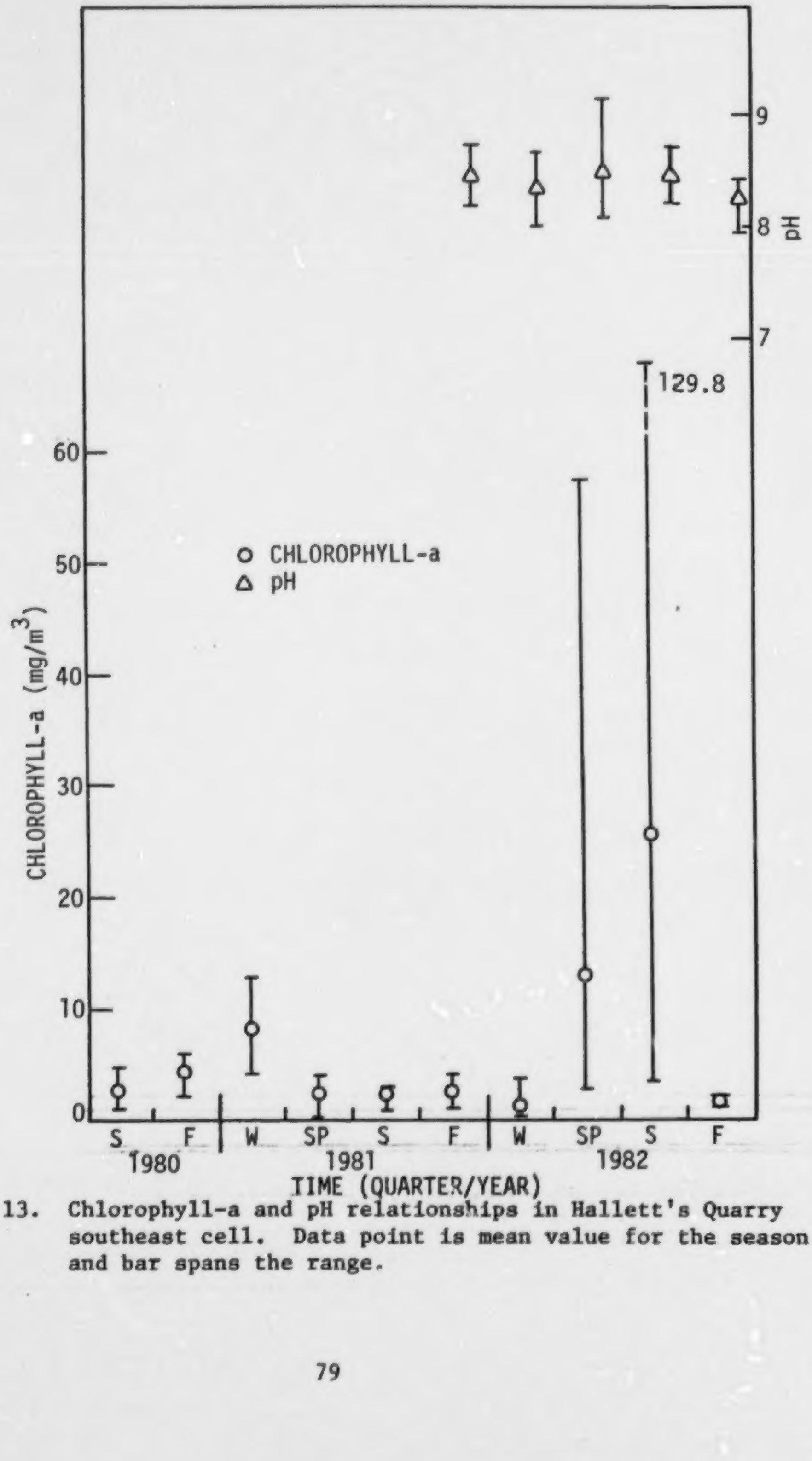


Fig. 13. Chlorophyll-a and pH relationships in Hallett's Quarry southeast cell. Data point is mean value for the season and bar spans the range.

TABLE 10. RAW WATER QUALITY PARAMETERS IN HALLETT'S QUARRY DURING PHASE I AND II (1981-1982) AND DURING PHASE III (1983) (mg/L UNLESS OTHERWISE NOTED)

Parameter	Phase I & II (1982)*			Phase III (1983)†		
	Avg.	σ‡	Range	Avg.	σ	Range
Alkalinity (as $CaCO_3$)	166	23	105-186	203	18	181-228
Total hardness (as $CaCO_3$)	297	22	233-323	358	23	328-393
Specific conductance (μmhos/cm)	590	44	474-669	686	55	662-791
Total dissolved solids	393	7.9	379-402	434	49	397-553
Suspended solids	6.1	6.2	0.5-30.6	8	5	4-19
Total PO_4	0.21	0.07	0.05-0.39	0.20	0.01	0.19-0.24
Ortho PO_4	0.12	0.10	0.04-0.44	0.08	0.05	0.01-0.15
NH_3 - N	0.23	0.15	0.04-0.63	0.63	0.53	0.15-1.62
NO_2 + NO_3 - N	1.88	1.70	0.19-4.91	6.20	0.25	5.74-6.40
Kjeldahl-N	0.55	0.33	0.25-1.92	0.91	0.60	0.49-2.10
Soluble SiO_2	13.1	2.34	9.1-15.6	13.5	0.96	11.8-14.4
COD	8.7	6.37	2.6-33.8	7.0	2.8	2.9-9.8
Chlorophyll-a (mg/m^3)	11.6	27.9	0.2-132.4	2.5	0.6	1.9-3.2

*Based on 28 observations except TDS (7 observations), Chlorophyll-a (48 observations).

†Based on 5 observations.

‡σ = standard deviation.

TABLE 11. 1981-83 CHLOROPHYLL A CONCENTRATIONS IN HALLETT'S QUARRY RAW WATER (mg/m^3)

Date	Chlor-a mg/m^3	Date	Chlor-a mg/m^3	Date	Chlor-a mg/m^3	Date	Chlor-a mg/m^3
1981							
11/9	2	3/3	0.9	7/12	28.2	11/15	0.7
11/23	1.8	3/10	0.4	7/20	4.6	11/17	0.7
12/2	2.6	3/15	1.8	7/26	132.4	11/29	1.1
12/7	3.2	3/17	1.1	7/28	130.7		
12/9	3.5	3/29	3.1	8/9	6.1	1983	
12/14	4	4/13	6.5	8/11	4.8	5/12	1.9
		4/26	59.5	8/18	2	5/16	1.2
1982		4/28	57.2	8/23	3.6	5/18	2.4
1/4	4	5/10	2.8	9/8	4.0	6/27	3.2
1/6	3.3	5/13	14.2	9/10	2.9	6/28	2.4
1/18	1.3	5/18	7.8	9/17	3.8	7/12	5.1
1/27	0.9	5/24	4.5	9/20	5.0		
2/1	0.8	6/7	7.1	9/30	4.2		
2/10	0.4	6/25	10	10/4	2.3		
2/17	0.2	6/28	3.5	10/18	2.6		
3/1	0.4	6/30	4	11/1	2.7		

During one of the algae blooms in summer 1982, two algae bioassays were conducted by David Millie, a graduate student in the Botany Department. The results of these two bioassays are given in Table 12, where the filter clogging algae are identified. Areal standard unit (asu) values were calculated by using the mean cell dimensions [31] except for the Volvox colonies and zygospores [65]. One asu is equal to 400 square microns. Using the standing crop and asu values, the late July total was 4818 asu/mL and the mid-August total was 1299 asu/mL. Due to the range of actual cell and colony size and the bioassay identification of some algae by genus only, the asu values were very rough estimates. Also, only the dominant taxa were given. Other algae were present but were not quantified due to their low concentrations.

TABLE 12. ALGAE BIOASSAYS

Dominant Taxa	Standing Crop	asu/Unit[*]
Late July, 1982 (total standing crop - 4330 cells or filaments/mL)		
Aphanizomenon flos-aquae	2913 filaments/mL[†]	1.38
Anabaena sp.[‡]	446 filaments/mL[†]	1.05
Oscillatoria sp.[‡]	207 filaments/mL	0.33
Chlamydomonas sp.	127 cells/mL	0.74
Mallomonas sp.	80 cells/mL	1.65
Mid August, 1982 (total standing crop - 1253 units/mL)		
Eudorina elegans	79 cells/mL	0.785
Dictyosphaerium pulchellum	375 cells/mL	0.11
Volvox globator	10 colonies/mL[§]	16
Volvox globator (zygospores)	40 zygospores/mL	16
Dinobrynon divergens[‡]	69 cells/mL	0.80
Aphanizomenon flos-aquae	206 flakes/mL[†]	1.38
Merismopedia tenuissima	69 colonies/mL[¶]	0.654

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[*] Values based on mean unit dimensions [31] except *Volvox* colonies and zygospores [65].

[†] 1 flake or filament is approx. equal to 10 cells.

[‡] Filter clogging [65,77].

[§] 1 *Volvox globator* colony contains as many as 17,000 cells.

[¶] 1 colony is 4 cells.

FILTRATION RESULTS, PHASE I

Phase I operation of the plant continued from Oct. 6, 1981 through Dec. 9, 1982 in order to obtain data during the full range of seasons. The winter of 1981-82 was unusually severe with extremely low temperatures and heavy snow. The snow melt runoff in the spring of 1982 and the heavy rains thereafter raised the water level higher than it had been in several years, and carried substantial plant nutrients into the quarry which resulted in a series of algal blooms. These blooms had a dramatic impact on the filtration results, as will be shown later.

Slow Sand Filtration Results, Phase I

The slow sand filter was operated at a rate of 0.12 m/h approach velocity (3.1 million gallons per acre per day). The maximum terminal head loss available to the overflow of the filter was 135 cm and filter runs were terminated when the overflow head loss was reached. The length of the filter runs varied substantially through the four seasons, as shown in Table 13. The best raw water occurred during the winter under the ice and this coincided with the longest runs of 123 days. The shortest runs at 9 days each occurred in the summer during periods of severe algal blooms.

Turbidity removal by slow sand filter--

In spite of the wide variation of run length associated with the varying raw water quality, the filtrate quality was consistently good. Typical results showing turbidity vs time for Runs A, B, F, G, J, and K of the slow sand filter are presented in Figs. 14 through 19. Note that the first two days of the run are shown on an expanded time scale to permit better illustration of the initial improvement period of the filter run.

The mean influent and effluent turbidity values for all the slow sand filter runs are presented in Table 14. The table presents the mean effluent values for two time periods, the first two days of the run and the remainder of the run. In calculating the mean values, the turbidity graphs were divided into nearly linear segments and the area under the graph was calculated to determine the mean. The number of values used in each calculation is shown in parentheses. Because Run B was long and had substantial trends in influent turbidity, the data for the run have been subdivided into four periods during the run. Similarly, in a few other runs, the first two days have been averaged separately from the remainder of the run.

The filtrate turbidity of the first two runs (A and B, Figs. 14 and 15) was higher than all subsequent runs. The turbidity of the later runs was consistently about 0.1 NTU except during the initial improvement period. The initial improvement period was less than two days' duration, except possibly for the first two runs.

The turbidity results of Run B (Fig. 15) are interesting because they show the gradual improvement in raw and filtrate quality after the ice formed on the quarry and then the sudden deterioration of raw water quality when snow melt runoff began reaching the quarry. During the transition, a ten- to

TABLE 13. DURATIONS AND INITIAL HEAD LOSSES OF SLOW SAND FILTER RUNS

Run Designation	Start	End	Duration* (days)	Initial Head Loss (cm)
A	10/19/81	11/22/81	34	12
B	12/01/81	04/03/82	123	11
C	04/07/82	04/29/82	22	18
D	05/05/82	05/18/82	13	18
E	05/25/82	06/06/82	12	18
F	06/10/82	06/19/82	9	18
G	06/23/82	07/14/82	22	18
H	07/21/82	07/31/82	10	17
I	08/04/82	08/24/82	20	18
J	09/15/82	10/26/82	41	19
K†	11/02/82	12/09/82	37‡	20

*Slow sand filter runs were terminated at overflow. Overflow was at a head loss of 136 cm.

†Filter surface was agitated in place rather than scraped prior to this run.

‡Run was ended prior to overflow at a head loss of 81 cm.

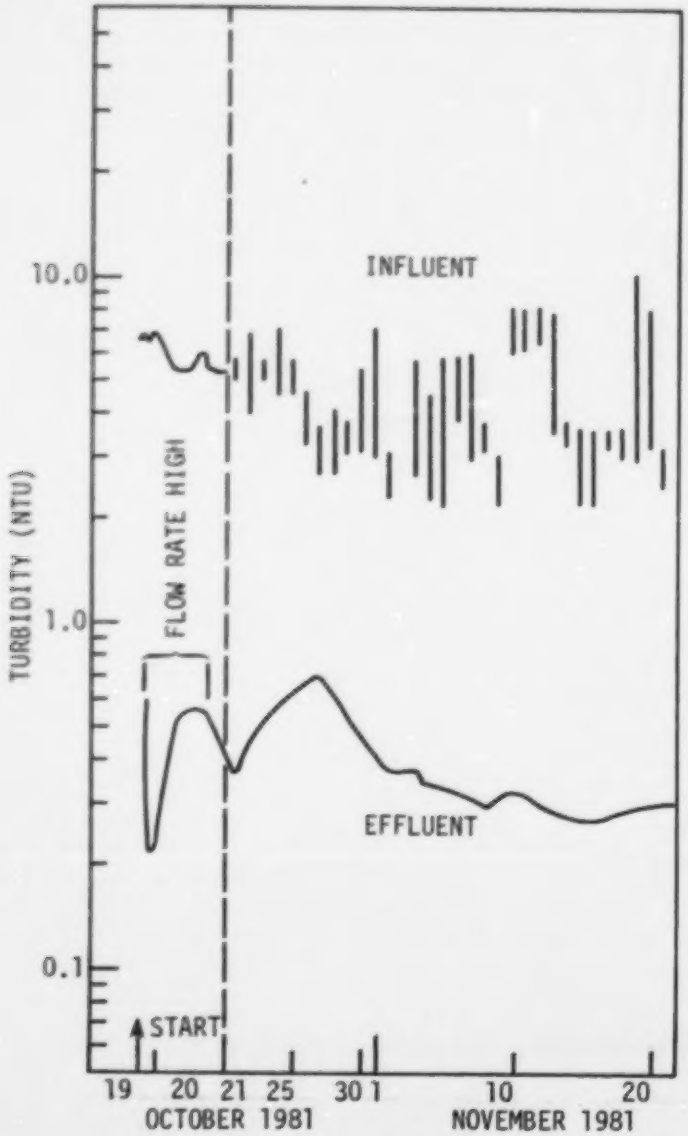

Fig. 14. Turbidity of influent and slow sand filter effluent during Run A. Influent bars show range of values recorded each day.

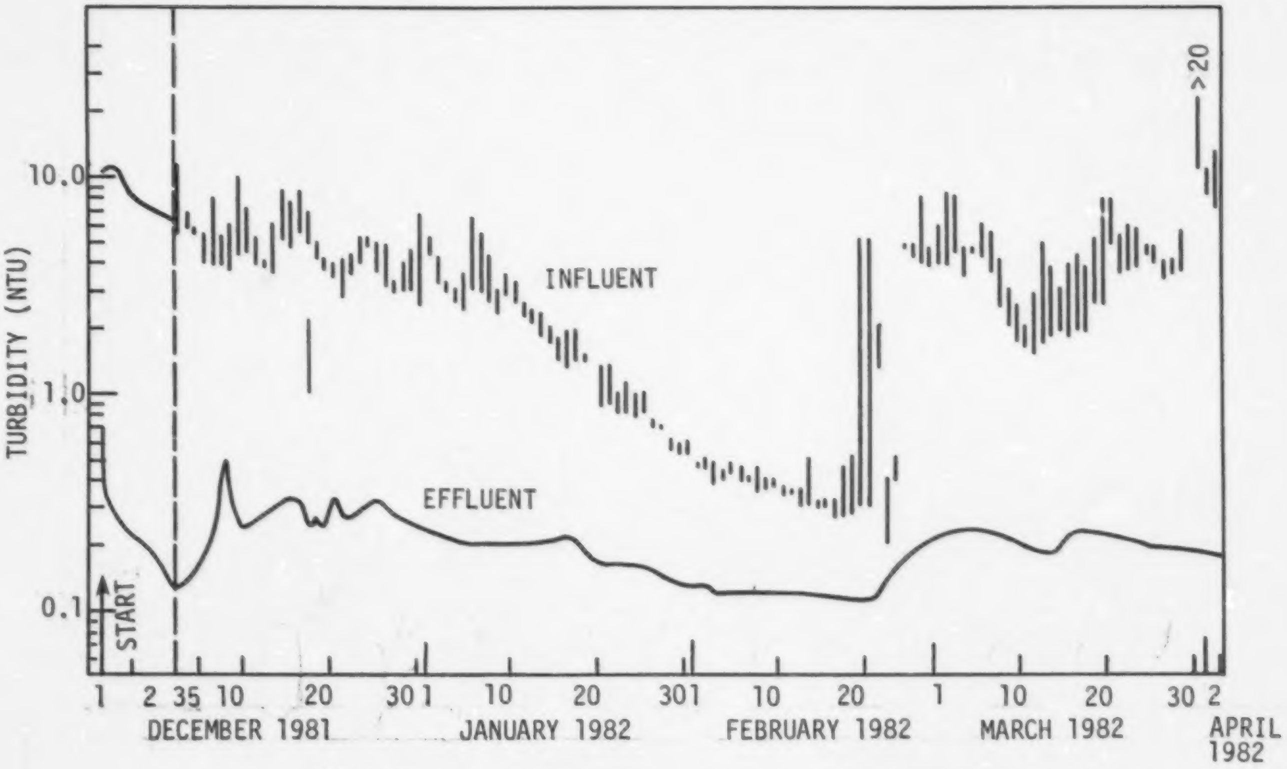

Fig. 15. Turbidity of influent and slow sand filter effluent during Run B. Influent bars show range of values recorded each day.

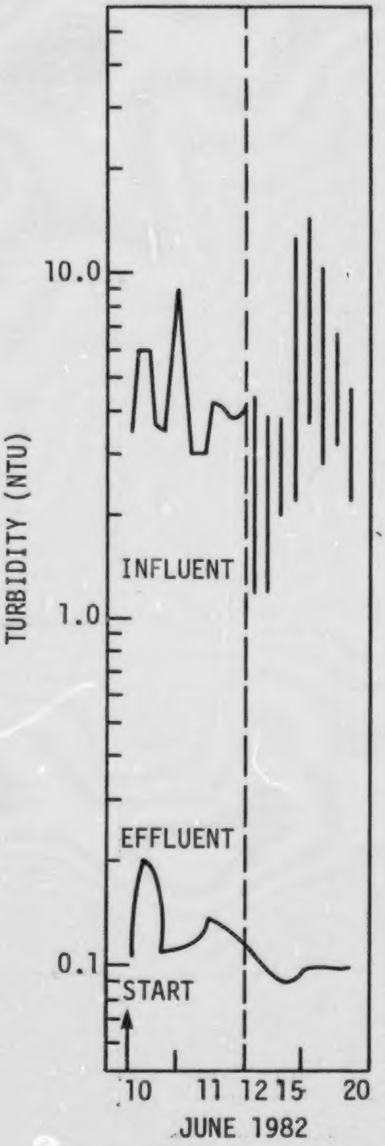

Fig. 16. Turbidity of influent and slow sand filter effluent during Run F. Influent bars show range of values recorded each day.

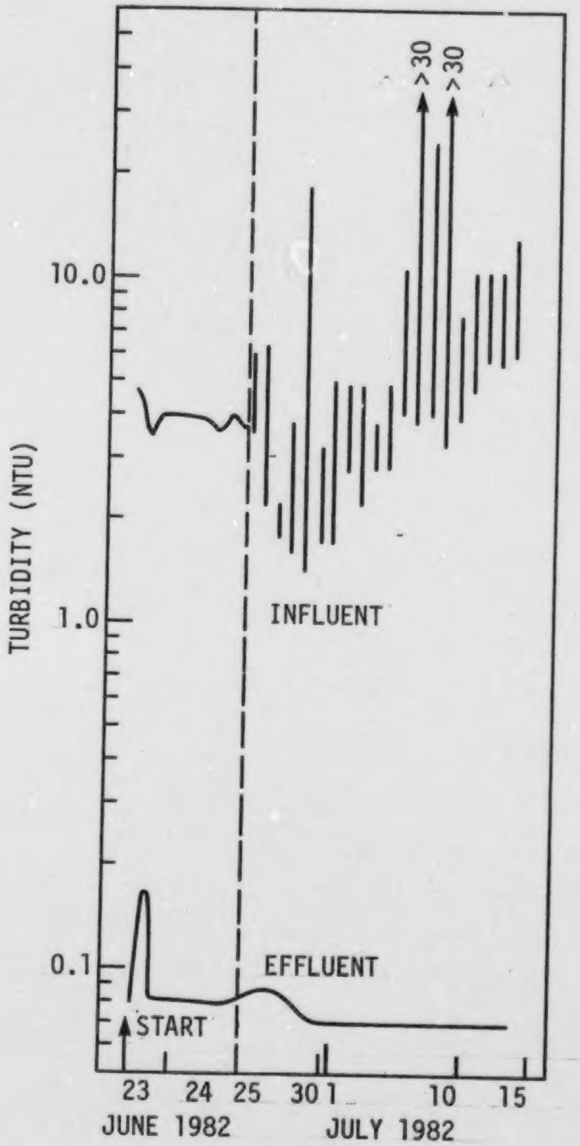

Fig. 17. Turbidity of influent and slow sand filter effluent during Run G. Influent bars show range of values recorded each day.

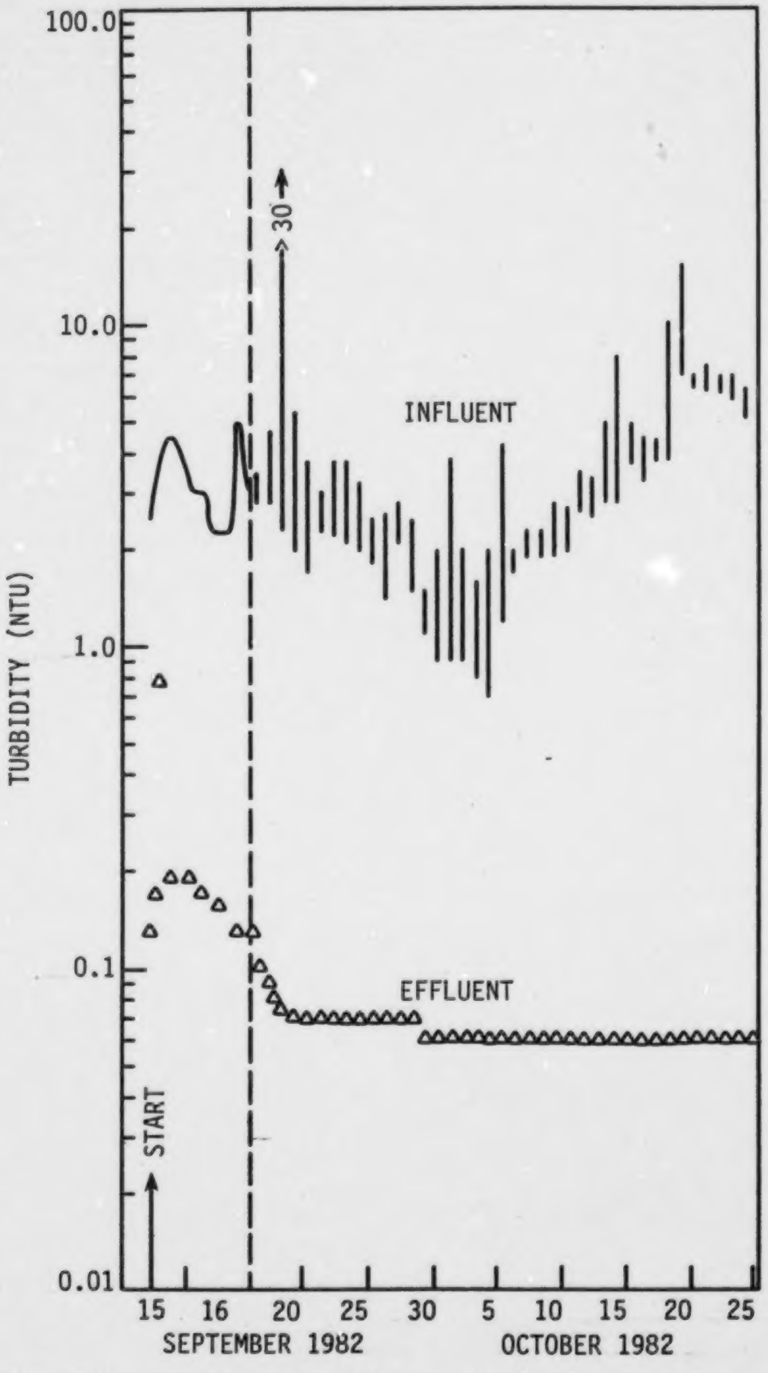

Fig. 18. **Turbidity of influent and slow sand filter effluent during Run J. Influent bars show range of values recorded each day.**

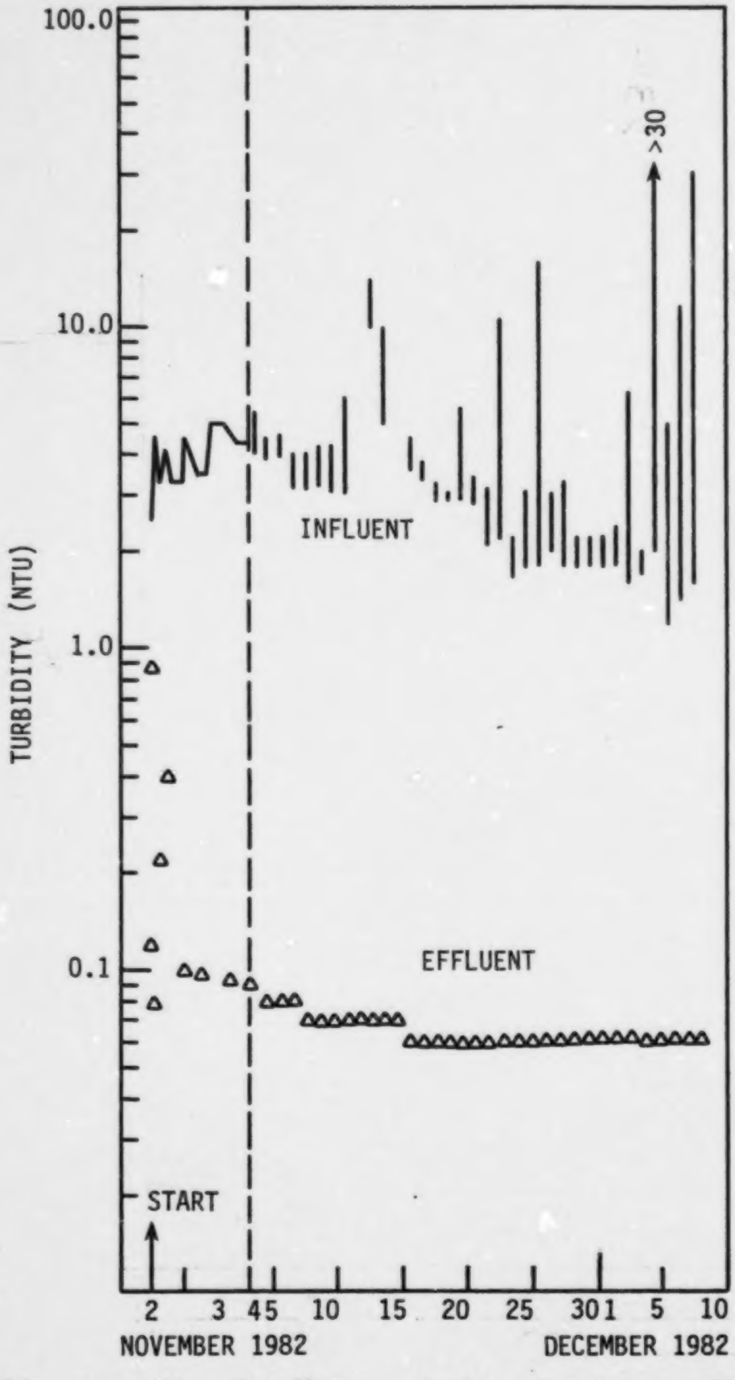

Fig. 19. Turbidity of influent and slow sand filter effluent during Run K. Influent bars show range of values recorded each day.

TABLE 14. TURBIDITY IN INFLUENT AND EFFLUENT OF SLOW SAND FILTER. REPORTED IN NEPHELOMETRIC TURBIDITY UNITS (NTU)

Run Number	Influent		Effluent					
			First Two Days			Remainder		
	Mean (NTU)	(Number)*	Mean	(Number)*	Percent Reduction	Mean	(Number)*	Percent Reduction
A	4.4	(33)	0.42	(2)	90.5	0.39	(32)	91.1
B	8.4	(2)	0.28	(2)	96.7			
B (12/3-16/81)	5.4	(14)				0.26	(14)	95.2
B (12/16/81-2/17/82)	2.2	(62)				0.20	(62)	90.9
B (2/17-4/3/82)	3.8	(43)				0.19	(43)	95.0
C	6.9	(23)	0.24	(2)	96.5	0.24	(21)	96.5
D	9.8	(2)	0.53	(2)	94.6			
D	2.8	(11)				0.23	(11)	91.8
E	3.0	(2)	0.32	(2)	89.3			
E	5.9	(11)				0.13	(11)	97.8
F	4.9	(10)	0.14	(2)	97.1	0.10	(8)	98.0
G	6.6	(22)	0.095	(2)	98.6	0.046	(20)	99.3
H	12	(2)	0.14	(2)	98.8			
H	19	(9)				0.14	(9)	99.3
I	7.4	(2)	0.46	(2)	93.8			
I	4.6	(19)				0.095	(19)	97.9
J	3.9	(42)	0.13	(2)	96.7	0.065	(40)	98.3
K	4.6	(36)	0.10	(2)	97.8	0.065	(36)	98.6

*Number of individual values used to calculate the mean value.

fifteen-fold increase in raw water quality resulted in only a two-fold increase in filtrate quality.

Particle removal by slow sand filter--

Typical particle count data for the same filter runs are shown in Figs. 20 through 24. These data look very much like the turbidity data, showing the same trends. Only a limited number of samples were taken for analysis each week, so that points are less numerous. Because of our interest in Giardia cyst-sized particles, the figures present the number of 7 to 12 µm sized particles per mL.

Particle count data in the 7-12 µm size range for all slow sand filter runs are summarized in Table 15. This table presents the mean influent and effluent counts during two periods, the initial two days of the run and the remainder of the run. The number of individual samples averaged is shown in parentheses.

Several things are evident in Table 15: (i) Removal in the first two days is worse than in the remainder of the run with the exception of Runs B and C; (ii) Low influent counts generally result in lower percent reductions than high influent counts (this is evident in Runs B and D); (iii) After Run D, the percent removal was consistently high, never below 96.9% and generally above 99%. Thus the filter improved over the series of runs.

Similar trends are evident in the data for total particle count in all 12 channels (1-60 µm size range) as presented for five filter runs in Figs. 25 through 28. The main difference is that the total number of particles per mL in the influent and effluent was from 1 to 2 logs higher than the corresponding number of 7-12 µm particles.

A summary of the 1-60 µm particle count data for all filter runs is presented in Table 16. Again the data are presented for the first 2 days of the runs, and for the remainder of the runs. The runs are presented in identical time periods as the prior turbidity data (Table 14) and 7-12 µm particle data (Table 15). Comparing Tables 15 and 16 with the data listed under the "remainder" heading, it is evident that high removals were achieved in both particle size ranges. Up through Run D, the percent reduction in total 1-60 µm particles was generally higher than the percent reduction in 7-12 µm particles. After Run D, the reverse is generally true although the differences are not large. It appears that longer service not only improved the removal of all particles, but did so preferentially for the particles in the 7-12 µm size range. After Run D, percent removal of total 1-60 µm particles was greater than 98% in all runs except Run I, and was usually higher than 99% (i.e., 2 log reduction). The lower percentage removals corresponded to low numbers in the influent water (Runs B, D, and I for example).

Comparing the turbidity data summarized in Table 14 with the 7-12 µm particle count data of Table 15, using the data under the heading "remainder," it can be concluded that after Run D, the percent reduction in 7-12 µm particles was higher than the percent reduction in turbidity. This

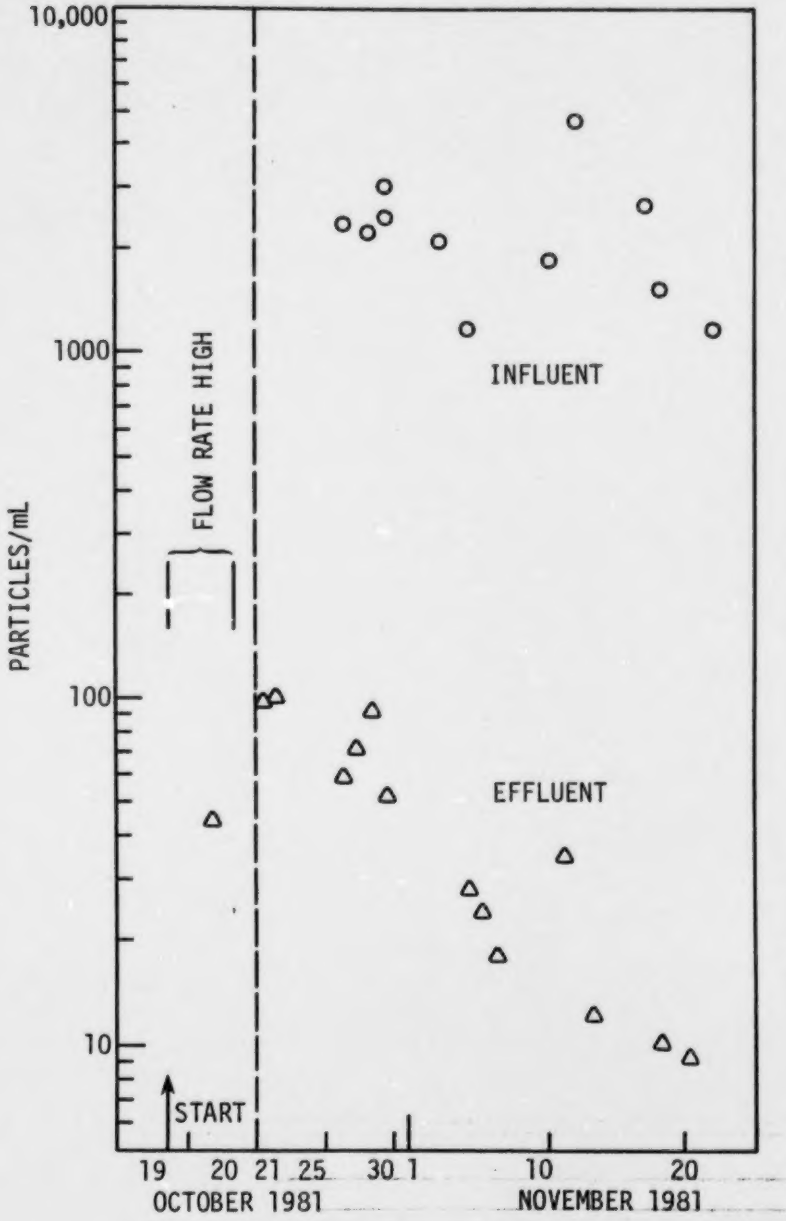

Fig. 20. Particle count in channels 7-9 (7- to 12-μm nominally) in the influent and effluent of slow sand filter during Run A.

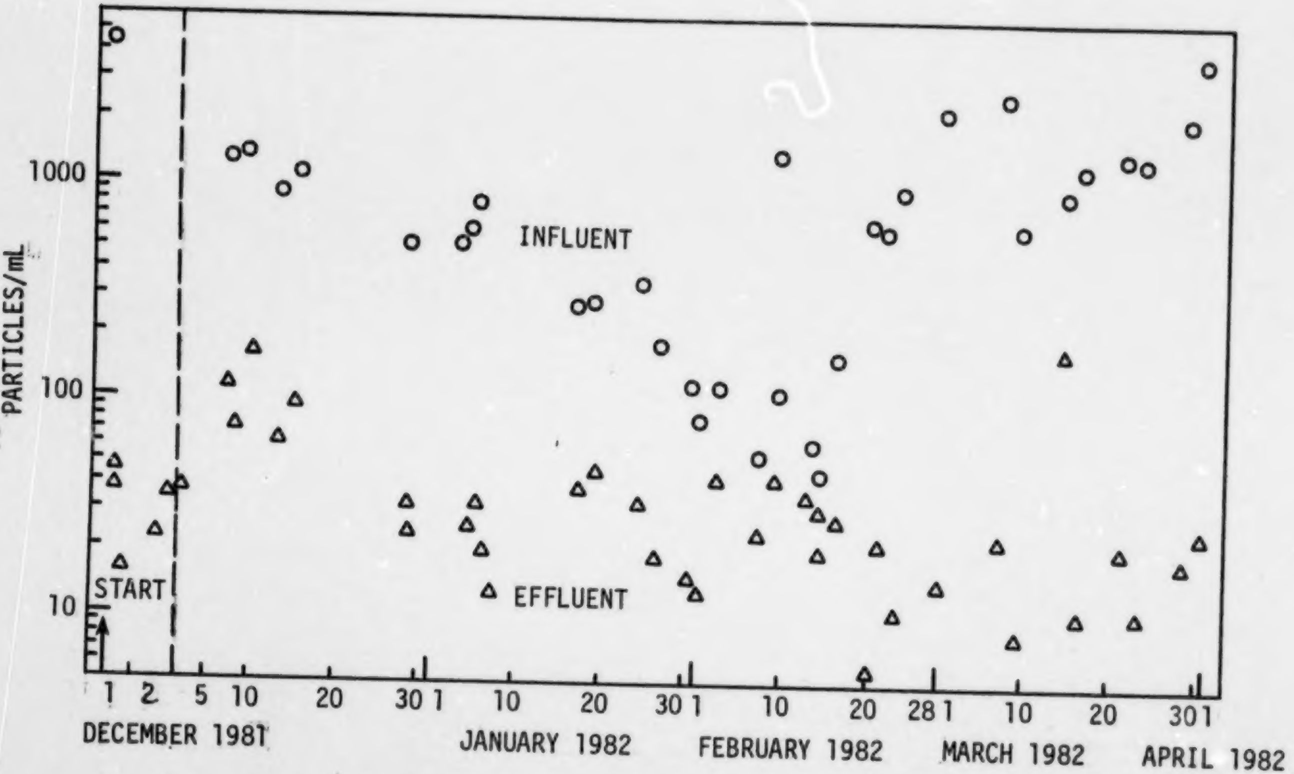

Fig. 21. Particle count in channels 7-9 (7-to 12-μm nominally) in the influent and effluent of slow sand filter during Run B.

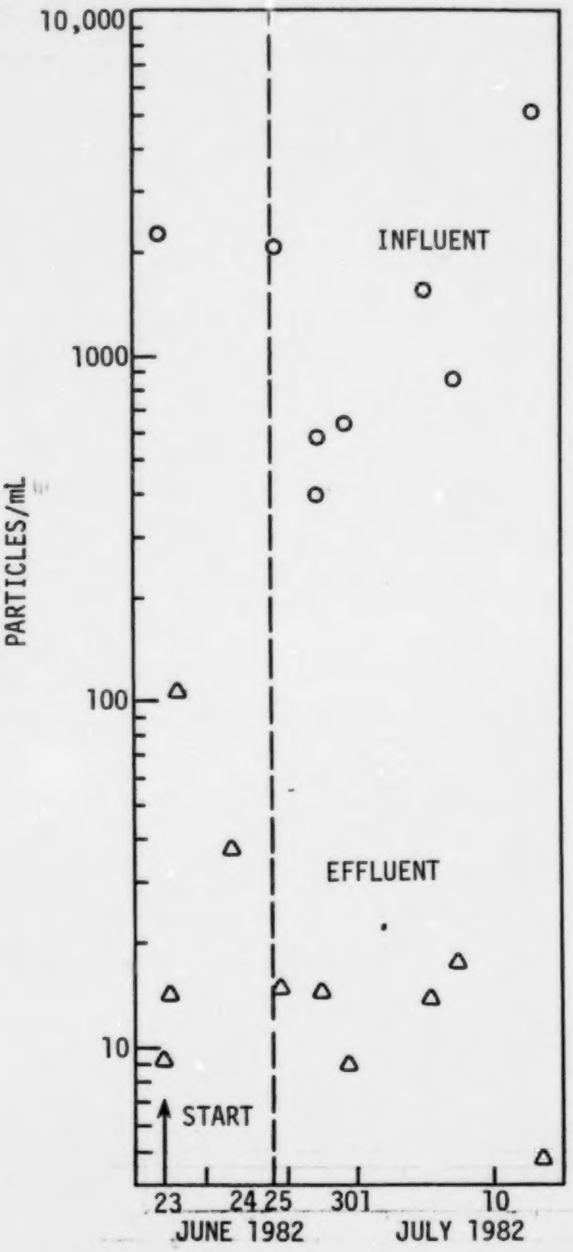

Fig. 22. Particle count in channels 7-9 (7- to 12-μm nominally) in the influent and effluent of slow sand filter during Run G.

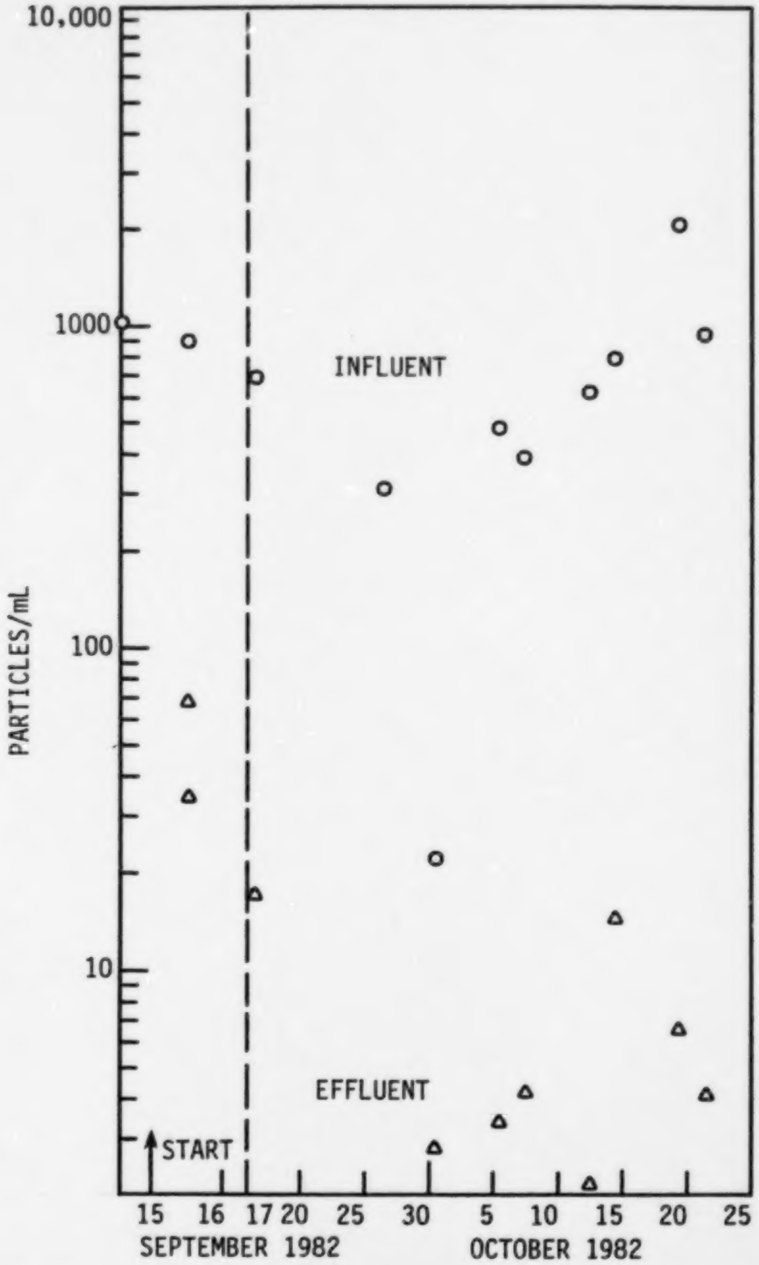

Fig. 23. Particle count in channels 7-9 (7- to 12-μm nominally) in the influent and effluent of slow sand filter during Run J.

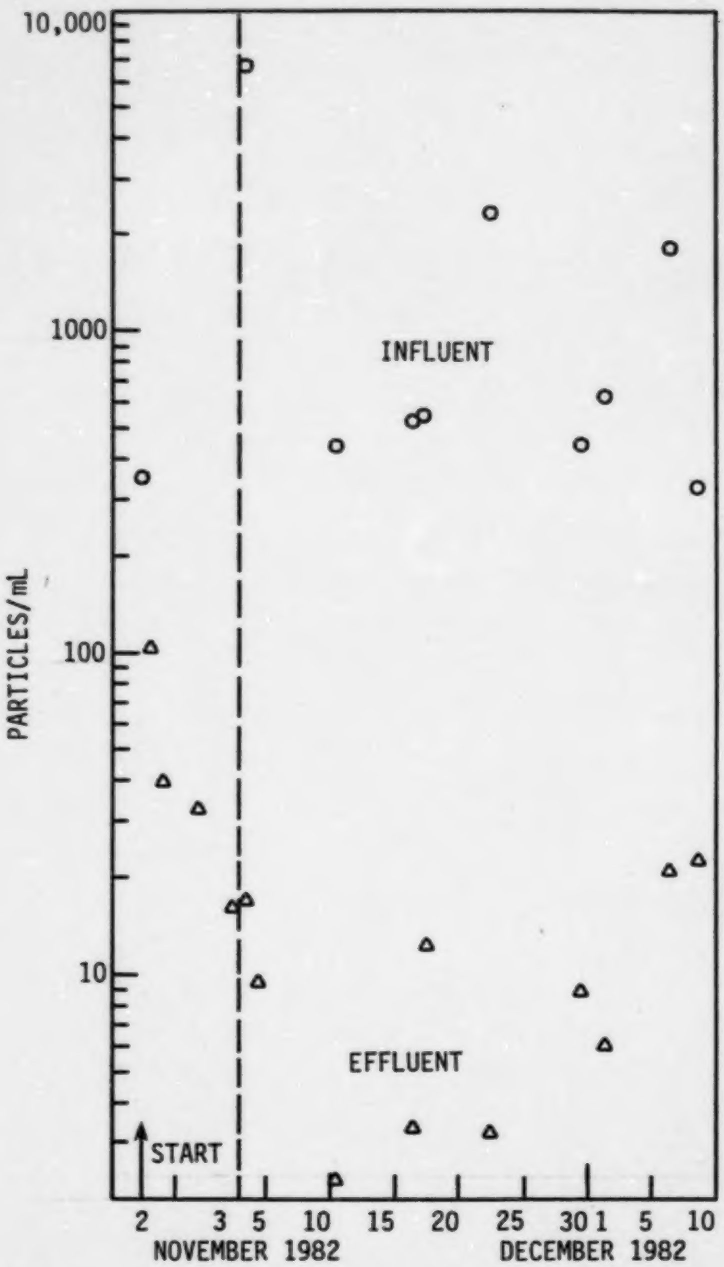

Fig. 24. Particle count in channels 7-9 (7- to 12-μm nominally) in the influent and effluent of slow sand filter during Run K.

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TABLE 15. PARTICLES PER mL IN INFLUENT AND EFFLUENT OF SLOW SAND FILTER IN 7-12 µm SIZE RANGE

			Effluent					
	Influent		First Two Days			Remainder		
Run Number	Mean	(Number)	Mean	(Number)*	Percent Reduction	Mean	(Number)*	Percent Reduction
A	2242	(10)	70	(2)	96.9	13	(12)	99.4
B	2857	(3)	33	(6)	98.8			
B (12/3-16/81)	1169	(4)				102	(5)	91.3
B (12/16/81-2/17/82)	265	(16)				29	(19)	89.1
B (2/17-4/3/82)	1534	(12)				15	(12)	99.0
C	3745	(7)	59	(6)	98.4	95	(6)	97.5
D	2412	(2)	104	(5)	95.7			
D	366	(2)				41	(2)	88.8
E	1425	(2)	69	(7)	95.2			
E	15732	(2)				46	(1)	99.7
F	10305	(2)	48	(4)	99.5	18	(2)	99.8
G	1704	(7)	37	(5)	97.8	12	(5)	99.3
H	6713	(1)	41	(3)	99.4			
H	20841	(2)				63	(2)	99.7
I	5460	(1)	412	(3)	92.4			
I	736	(4)				23	(4)	96.9
J	753	(10)	34	(6)	95.5	5	(8)	99.3
K	908	(10)	47	(4)	94.8	11	(10)	98.8

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*Number of individual values used to calculate the mean value.

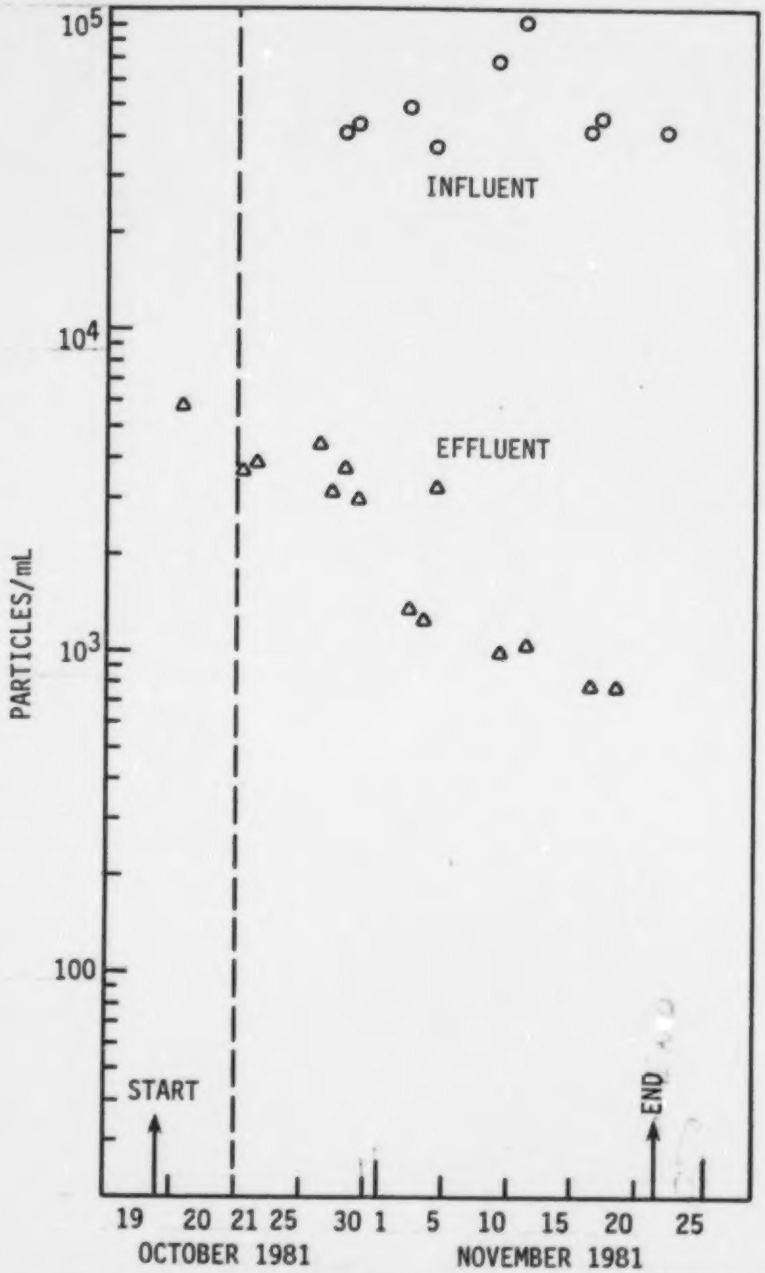


Fig. 25. Total particle count (1-60 µm) in the influent and effluent of slow sand filter in Run A.

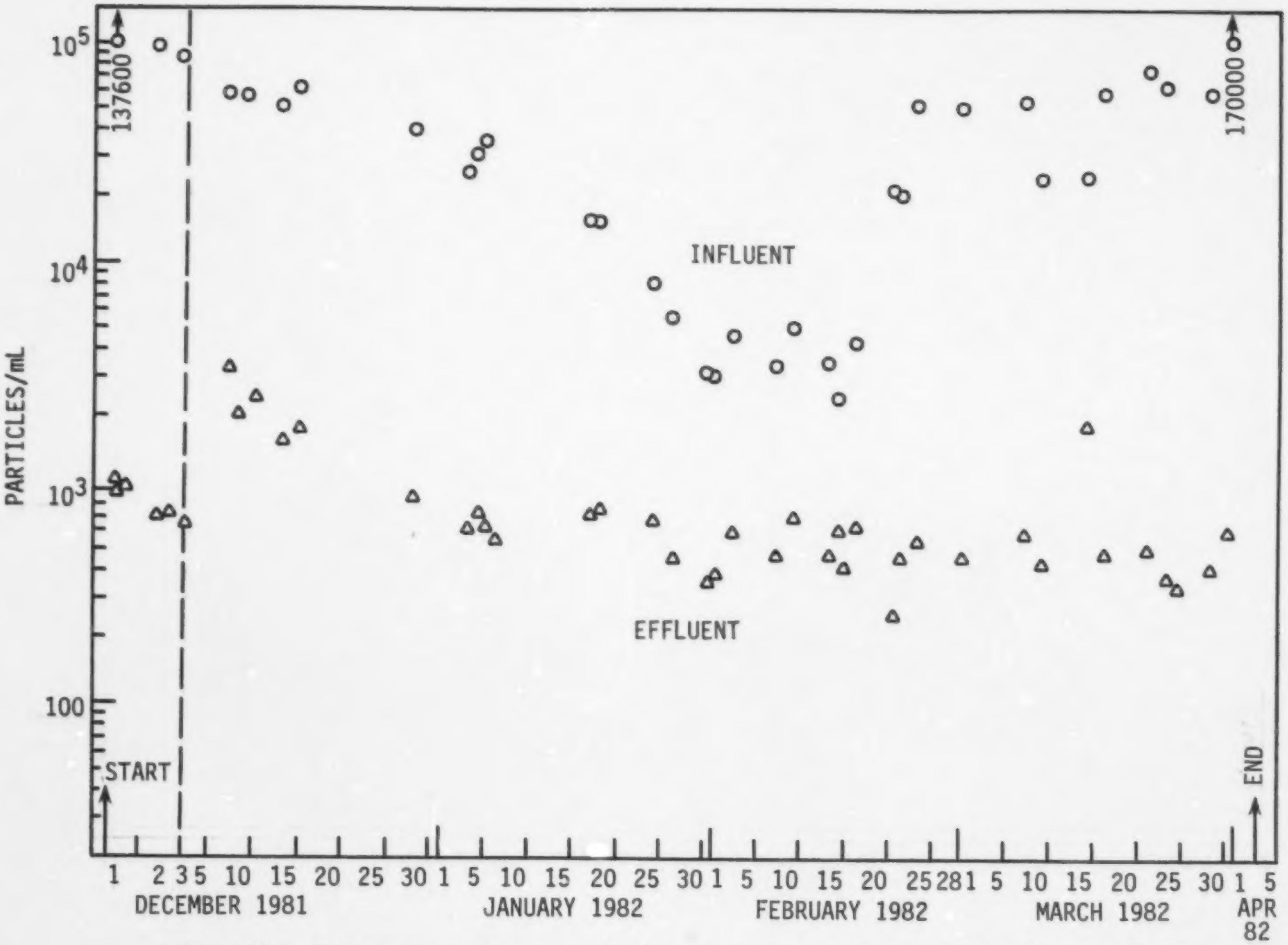


Fig. 26. Total particle count (1-60 μm) in the influent and effluent of slow sand filter in Run B.

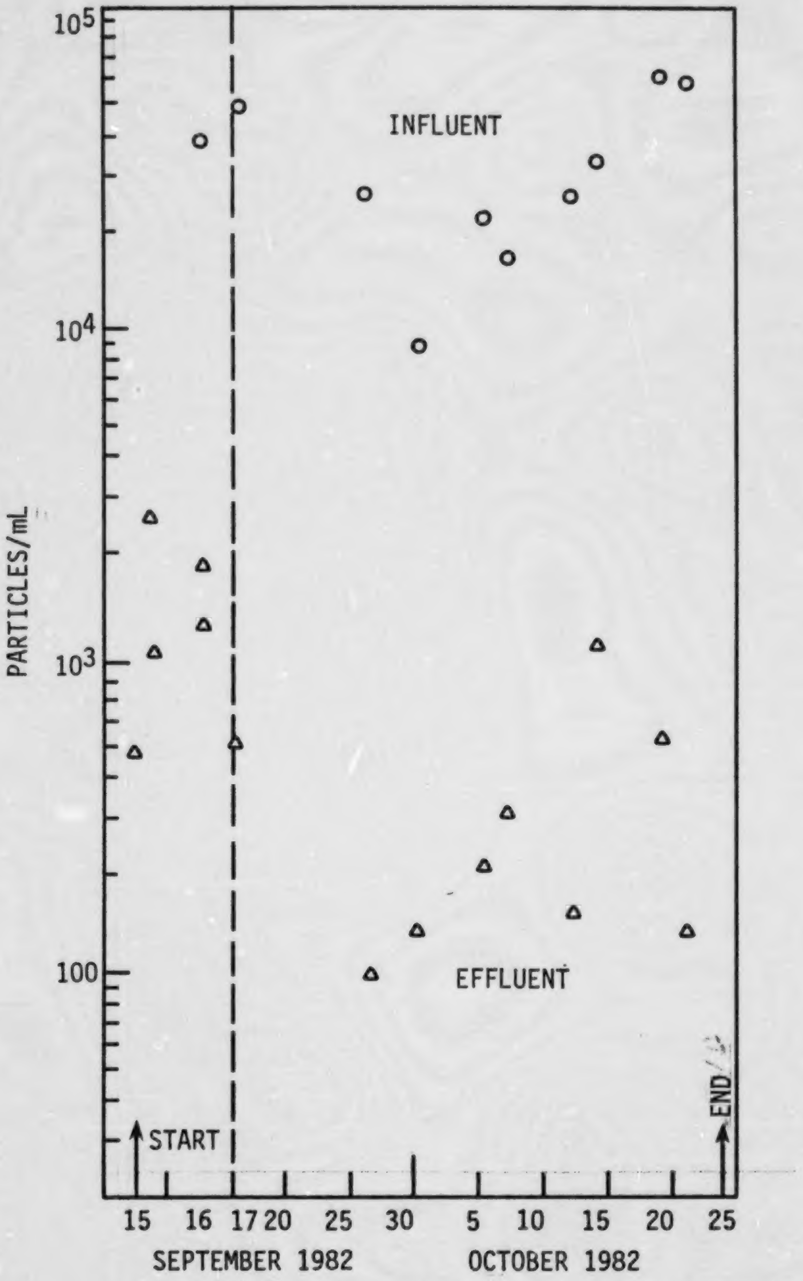


Fig. 27. Total particle count (1-60 μm) in the influent and effluent of slow sand filter in Run J.

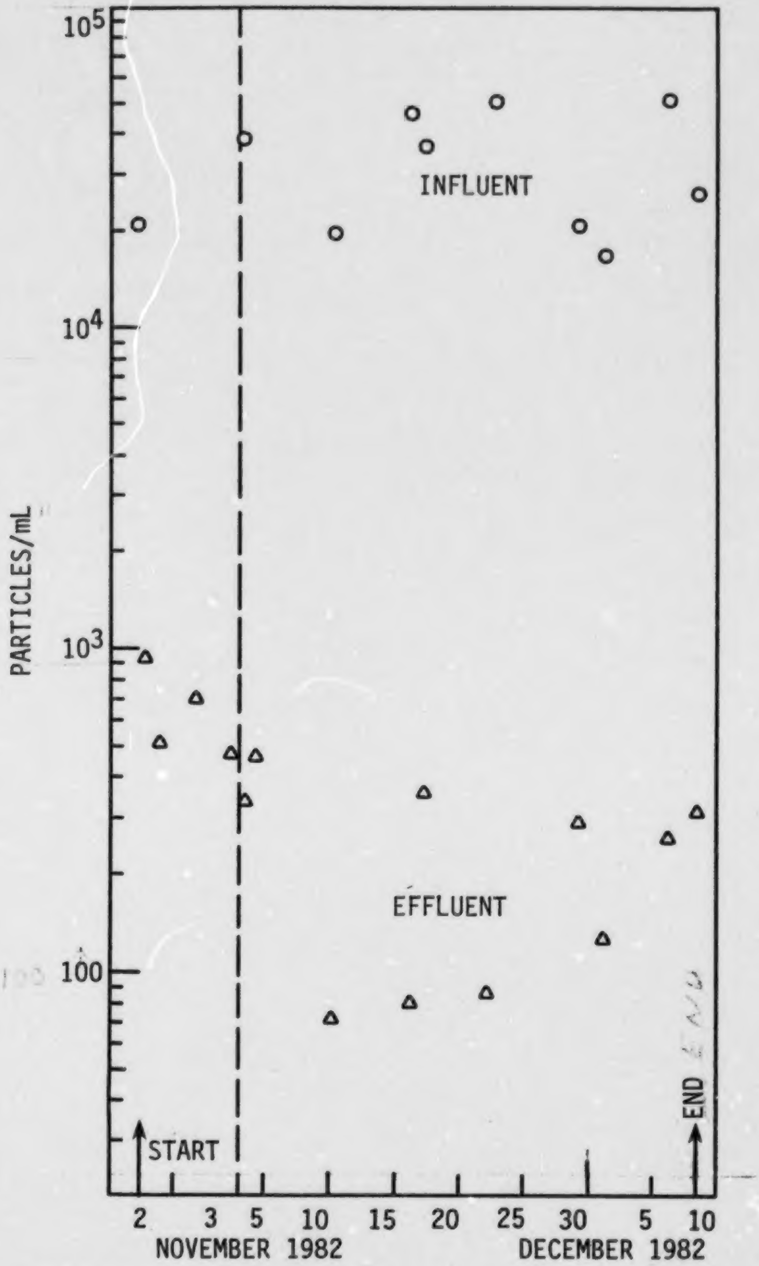


Fig. 28. Total particle count (1-60 μm) in the influent and effluent of slow sand filter in Run K.

TABLE 16. PARTICLES PER mL IN INFLUENT AND EFFLUENT OF SLOW SAND FILTER IN 1-60 μm SIZE RANGE

			Effluent					
	Influent		First Two Days			Remainder		
Run Number	Mean	(Number)	Mean	(Number)*	Percent Re-duction	Mean	(Number)*	Percent Re-duction
A	50,156	(10)	5740	(1)	88.6	2384	(13)	95.2
B	117,282	(2)	875	(5)	99.3			
B (12/3-16/81)	62,792	(5)				1971	(6)	96.9
B (12/16/81-2/17/82)	13,046	(16)				618	(19)	95.3
B (2/17-4/3/82)	56,182	(12)				566	(13)	99.0
C	96,252	(7)	1030	(6)	98.9	879	(5)	99.1
D	67,518	(2)	11,368	(5)	83.2			
D	9,439	(2)				687	(2)	92.7
E	21,722	(1)	7496	(6)	65.5			
E	32,854	(3)				630	(2)	98.1
F	42,400	(2)	866	(4)	98.0	298	(2)	99.3
G	37,723	(7)	617	(4)	98.4	226	(6)	99.4
H	197,114	(1)	342	(2)	99.8			
H	76,130	(2)				538	(2)	99.3
I	60,670	(1)	7130	(3)	88.2			
I	20,376	(4)				1466	(4)	92.8
J	33,355	(10)	1351	(5)	95.9	346	(9)	99.0
K	32,696	(10)	648	(4)	98.0	236	(10)	99.3

*Number of individual values used to calculate the mean value.

observation may be due in part to the fact that turbidity is not a linear function of particle number, but also because the effluent turbidity values were near the lower limit of the meter. Nevertheless, the important point is that turbidity would be a conservative predictor of performance where Giardia cysts were of concern.

Total coliform bacteria removal by slow sand filter--

Total coliform data were less consistent than the turbidity or particle count data. This was due to an inconsistent influent level obtained by spiking the influent with sewage. Typical results for the slow sand filter are shown in Figs. 29 through 33 for the same filter runs.

Mean influent and effluent levels of total coliforms are summarized for all runs in Table 17. Mean effluent levels are divided into two time periods, the first two days of the run and the remainder of the run. This table shows trends similar to the particle count data of Tables 15 and 16. In Table 17, the first two days are always poorer than the remainder of the run. There is a trend for better removal over the series of runs. Runs F through K are above 99% removal (except during the first two days). Total coliform removals are similar, but generally slightly higher than particle removals shown in Tables 15 and 16.

The standard plate count data were very erratic during the entire study. It is assumed that some bacteria were propagating in the effluent piping flow meter and turbidimeter and were released into the filtrate in an unpredictable pattern. In some cases, the effluent standard plate count exceeded the influent. Because of these difficulties, the data are not presented.

Head loss development for slow sand filter--

Head loss development for the slow sand filters always followed an exponential pattern which is typical of cake filtration (i.e., filtration in the schmutzdecke in this case). Head loss curves for four runs are presented in Figs. 34 through 36. These figures include the longest filter run (Run B, Fig. 34), the shortest run (Run F, Fig. 35), and the last run, which was preceded by raking rather than scraping (Run K, Fig. 36).

Because of the steeply rising head loss at the end of the filter runs, the run lengths could not be extended appreciably by increasing the terminal head loss available to the slow sand filter.

Algae removal by slow sand filter--

At periodic intervals, approximately every two weeks, samples were collected for chlorophyll analysis of the raw and filtered water. These data for chlorophyll-a have been summarized for each filter run of the slow sand filter in Table 18. The table lists the highest reported value during the filter run, the lowest reported value, and the mean of all values during the filter run. If only one sample was taken during a filter run, it is reported in the mean column.

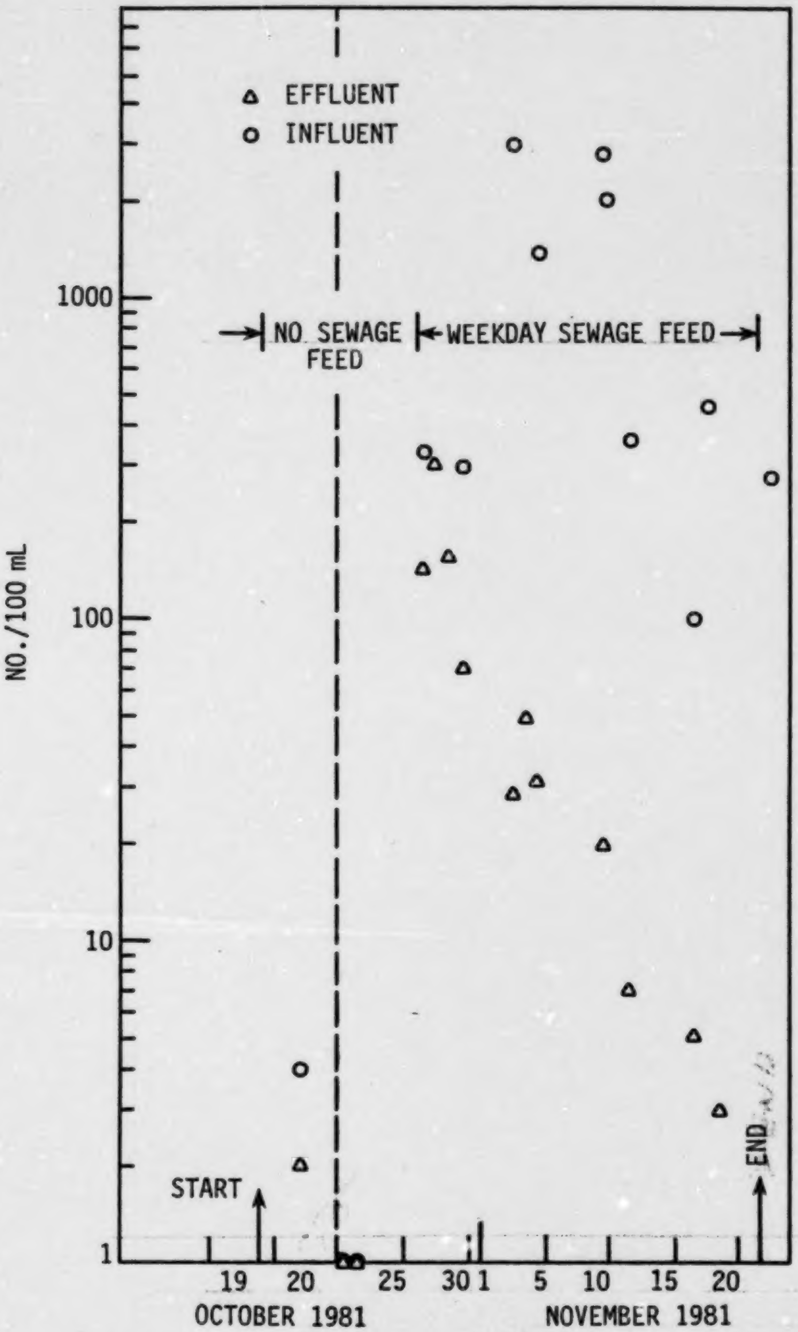


Fig. 29. Total coliform bacteria in influent and effluent of slow sand filter during Run A.

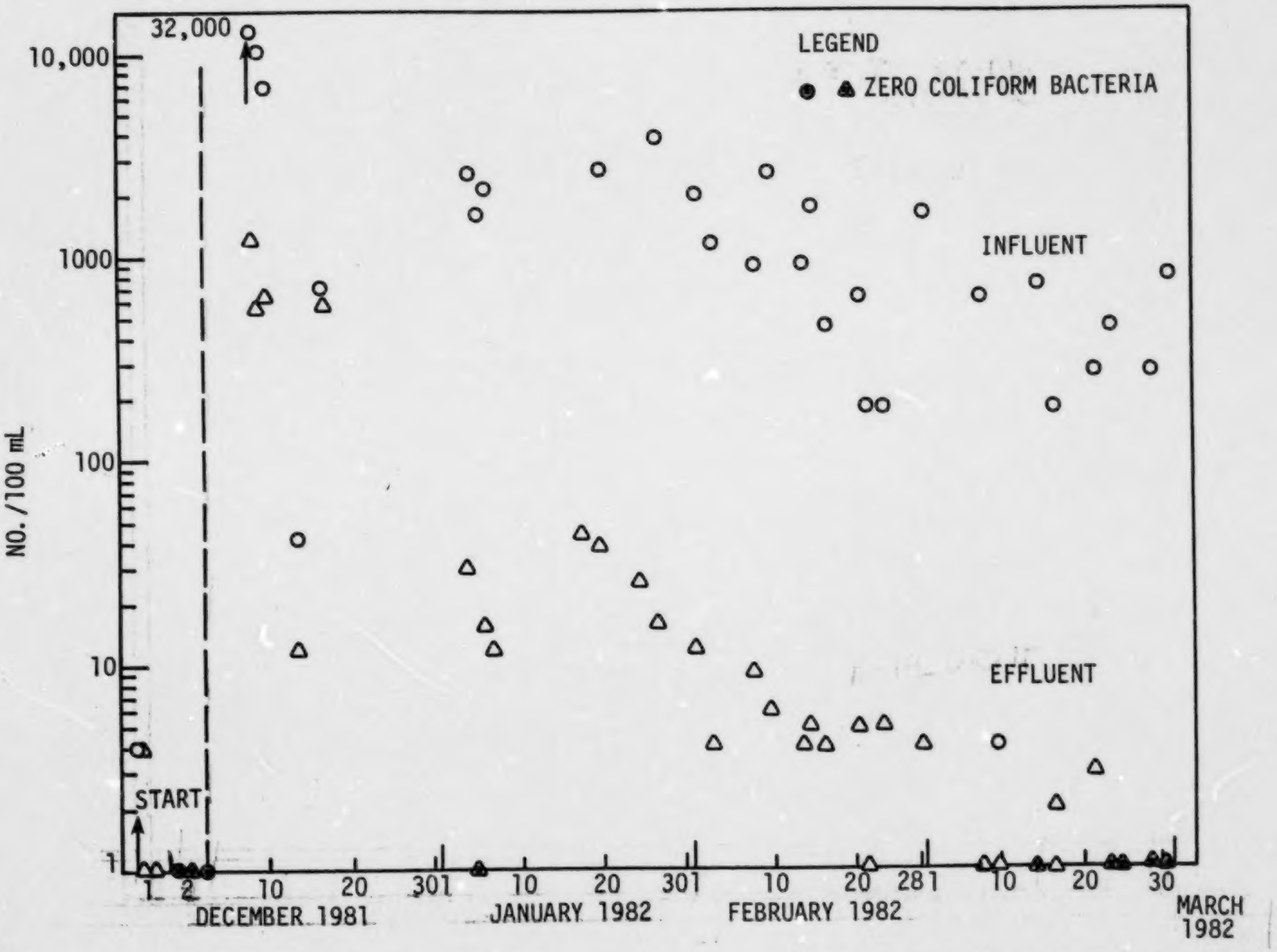


Fig. 30. Total coliform bacteria of influent and effluent of slow sand filter during Run B. Samples reported with zero coliform bacteria are plotted at the level of 1/100 mL.

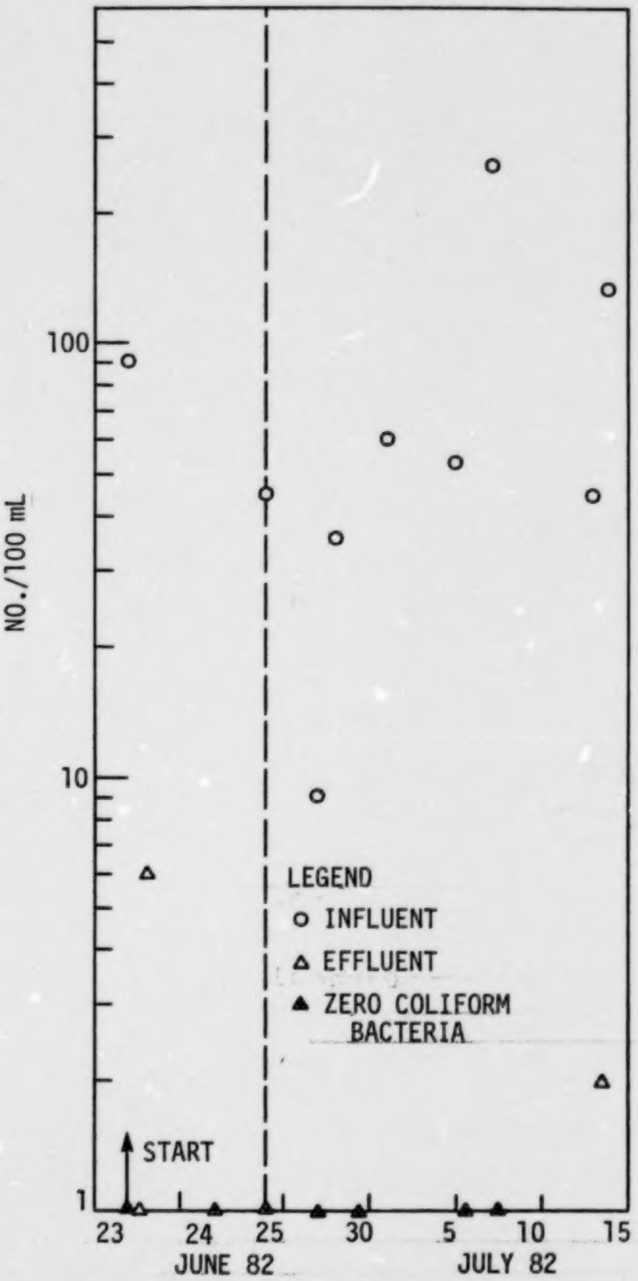


Fig. 31. Total coliform bacteria of influent and effluent of slow sand filter during Run G. Samples reported with zero coliform bacteria are plotted at the level of 1/100 mL.

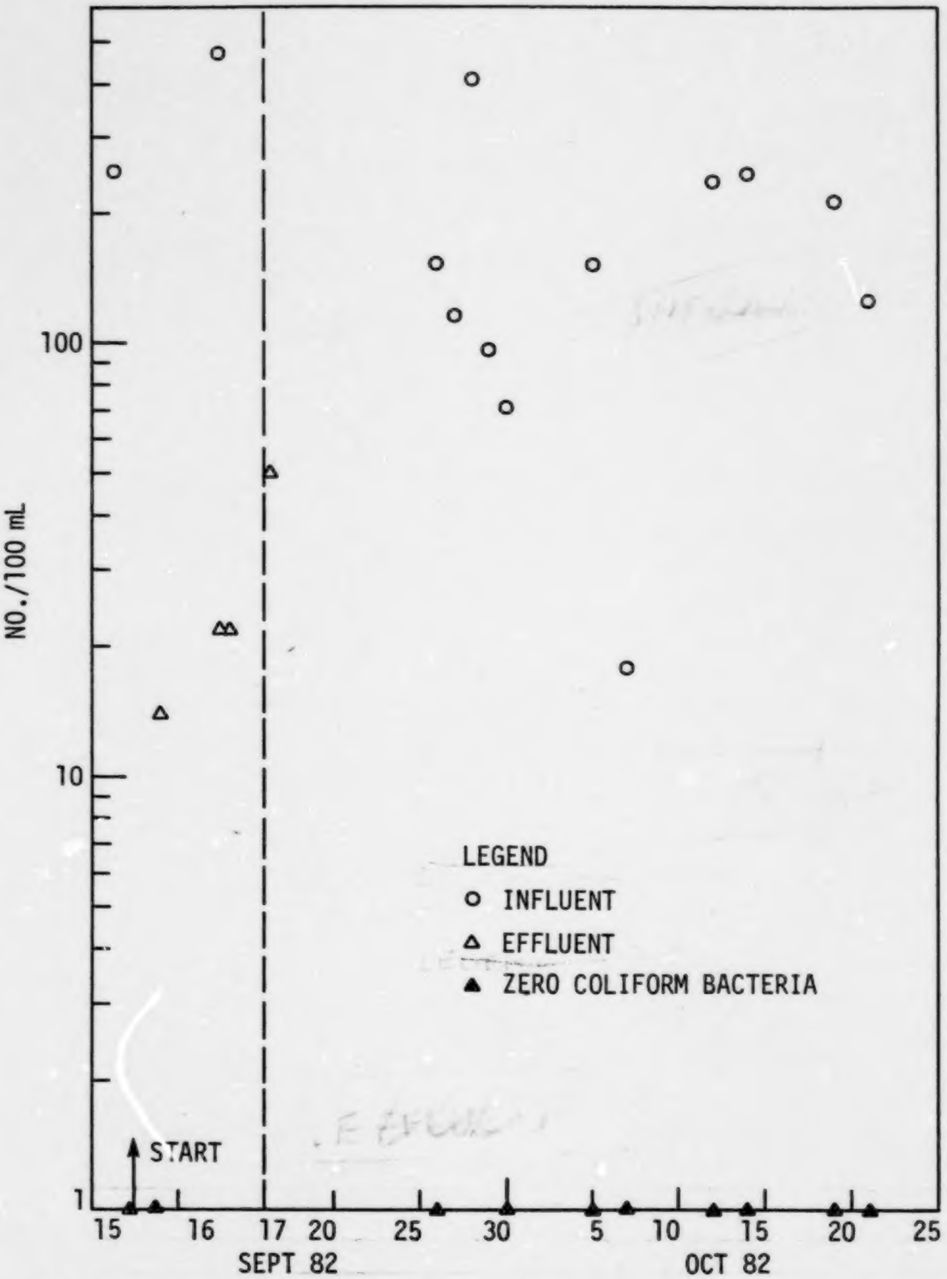


Fig. 32. Total coliform bacteria of influent and effluent of slow sand filter during Run J. Samples reported with zero coliform bacteria are plotted at the level of 1/100 mL.

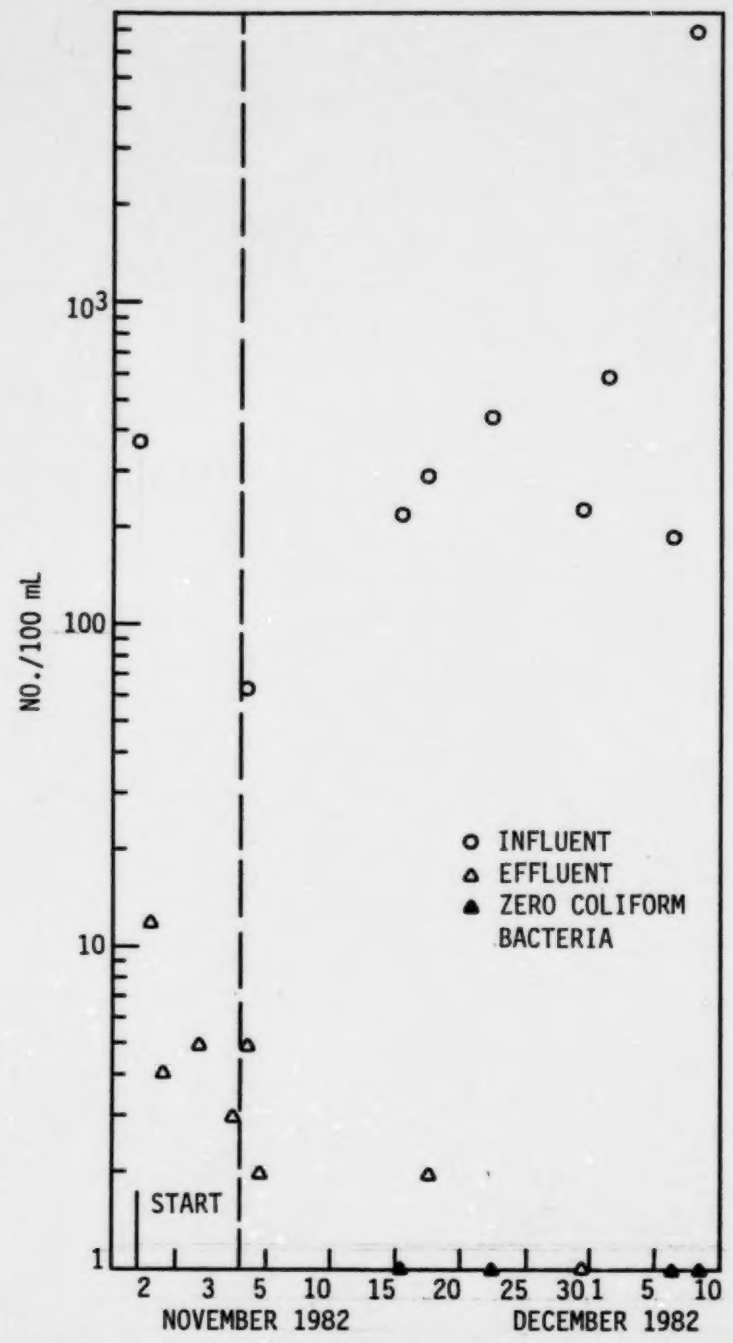


Fig. 33. Total coliform bacteria in influent and effluent of slow sand filter in Run K. Samples with zero coliform bacteria are plotted at the level of 1/100 mL.

TABLE 17. TOTAL COLIFORM BACTERIA PER 100 mL IN INFLUENT AND EFFLUENT OF SLOW SAND FILTER

Run Number	Influent		Effluent					
			First Two Days			Remainder		
	Mean	(Number)*	Mean	(Number)*	Percent Reduction	Mean	(Number)*	Percent Reduction
A	2032	(12)		No data		150	(14)	92.6
B	0	(3)	2.2	(3)	---			
B (12/3-16/81)	9791	(5)				494	(5)	95.0
B (12/16/81-2/17/82)	2119	(15)				15	(16)	99.3
B (2/17-4/3)	523	(12)				1.7	(13)	99.7
C	737	(6)	1.7	(6)	99.8	1.0	(4)	99.9
D	47	(4)	8.4	(5)	82.1	1.3	(2)	97.2
E	584	(4)	12	(5)	97.9	3.7	(3)	99.4
F	191	(2)	14	(5)	92.7	0.3	(2)	99.8
G	74	(8)	1.4	(5)	98.1	0.4	(5)	99.5
H	42	(3)	2	(2)	95.2		Bad data	
I	122	(3)	10	(4)	91.8	0.5	(2)	99.6
J	194	(12)	18	(6)	90.7	0	(8)	100.0
K	1044	(9)	6	(5)	99.4	0.7	(7)	99.9

*Number of individual values used to calculate the mean value.

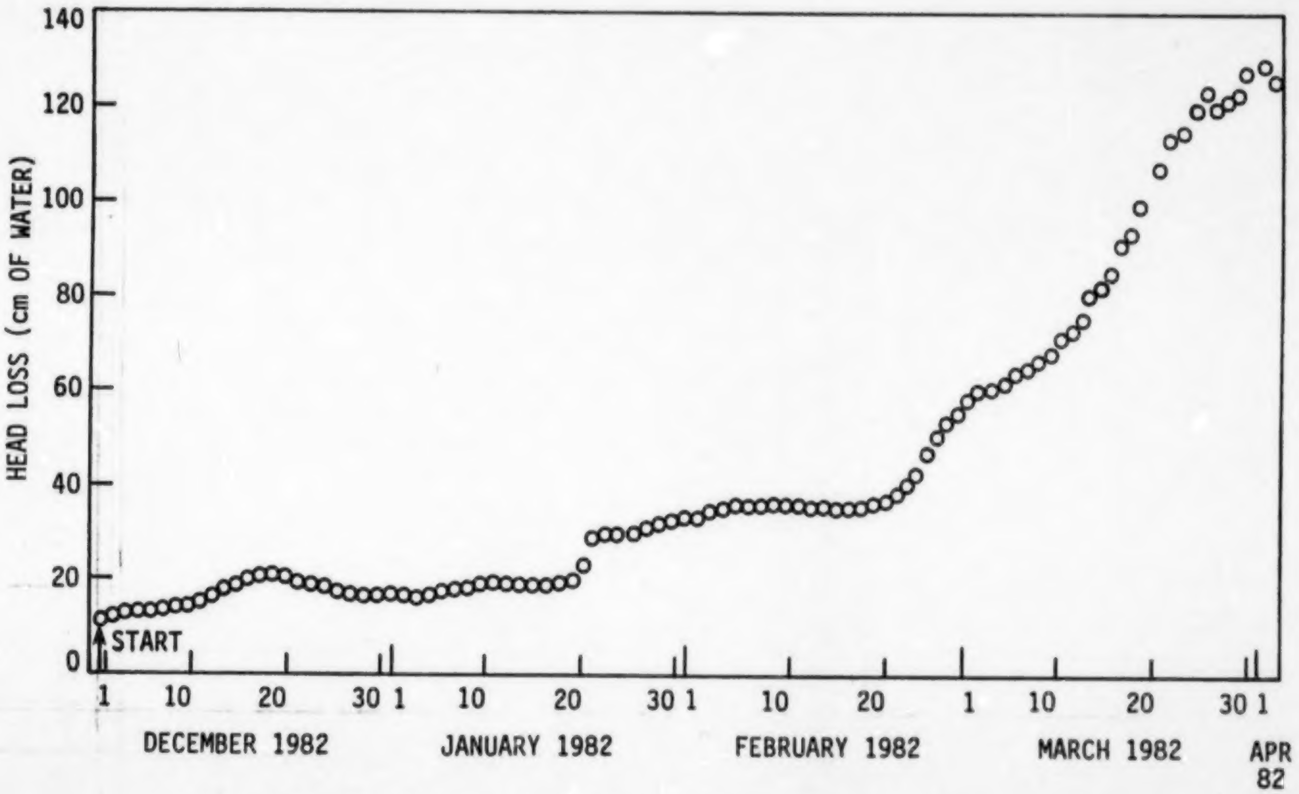


Fig. 34. Head loss development during Run B for slow sand filter.

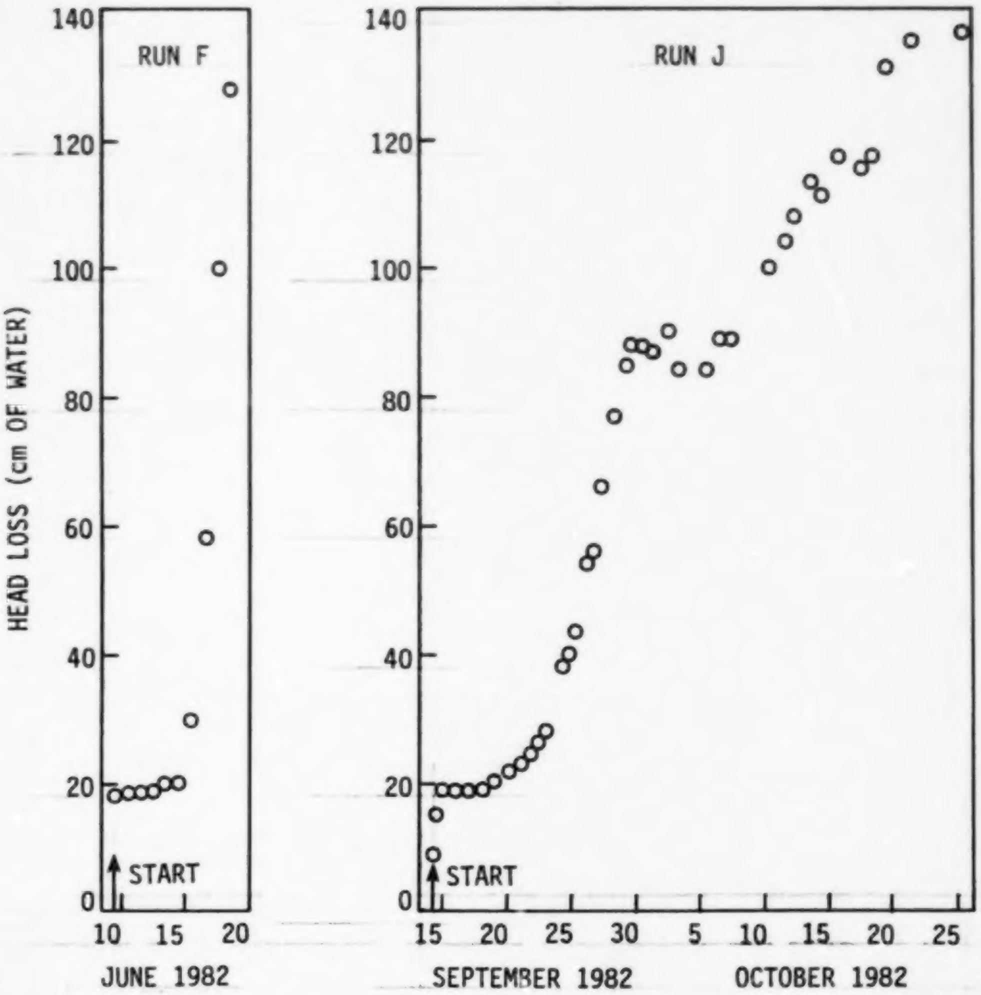


Fig. 35. Head loss development during Runs F and J for slow sand filter.

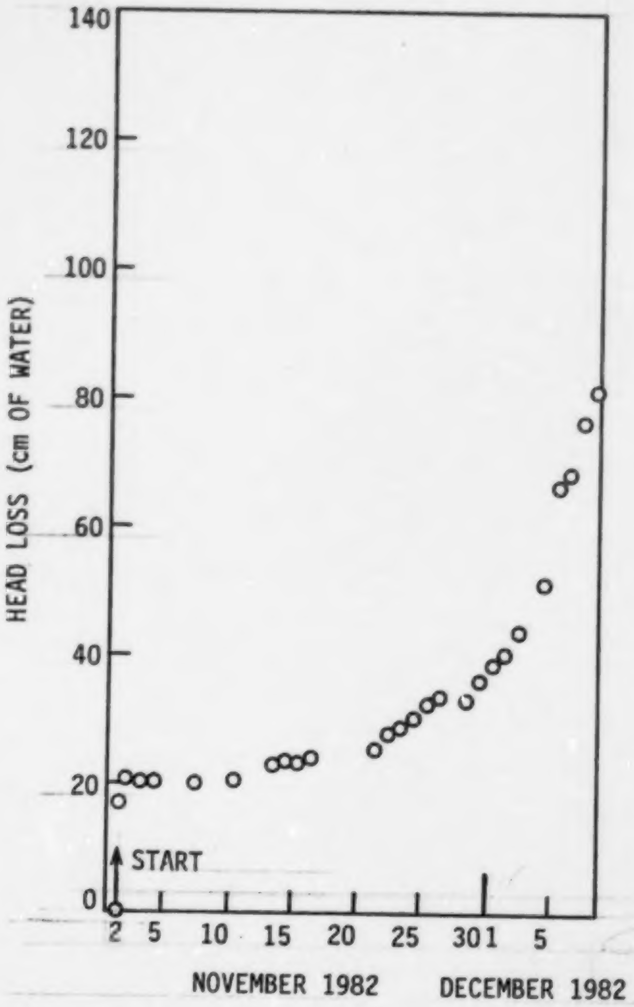


Fig. 36. Head loss development during Run K for slow sand filter.

TABLE 18. RAW WATER AND SLOW SAND FILTER EFFLUENT CHLOROPHYLL-A CONCENTRATIONS DURING PHASE I

Run Number	Raw Water				Slow Sand Filter Effluent				Percent Reduction
	High mg/m^3	Low mg/m^3	Mean mg/m^3	(Number)*	High mg/m^3	Low mg/m^3	Mean mg/m^3	(Number)*	
A			2.0	(1)			No Data		
B	4.0	0.2	1.9	(16)	1.8	0.0	0.35	(8)	81.6
C	60	6.5	41	(3)			0.59	(1)	98.6
D	14	2.8	8.5	(2)			0.0	(1)	100
E			No Data				No Data		
F			No Data				No Data		
G	28	3.5	11	(4)	0.0	0.0	0.0	(2)	100
H	143	132	138	(2)			0.20	(1)	99.9
I	6.1	2.0	4.1	(4)	0.0	0.0	0.0	(2)	100
J	5.0	2.3	3.6	(5)	0.7	0.0	0.18	(4)	95.0
K	1.1	0.70	0.90	(2)			Bad Data		

---

*Number of individual values used to calculate the mean value.

It is evident that the algae removal by the slow sand filter was excellent and improved over the series of filter runs. The algae removal was over 95% beginning with Run C and was 3 log reduction or better in Runs G, H, and I.

Oxygen utilization in slow sand filtration--

Periodic samples of influent and filter effluent water were analyzed for dissolved oxygen in the field. Table 19 summarizes the data. Since relatively few samples were collected, they are presented in three columns: the first sample of a filter run, the last sample of the filter run, and all the remaining samples in between. The purpose was to show whether there was an increasing usage as a filter run progressed. If only two samples were collected they are presented as the first and last sample. If only one sample was collected it is presented in the "mean" column. DO was present in all samples collected.

In general, Table 19 supports the expectation that DO utilization by the organisms in the filter increases as the run progresses. This was evident in all runs except I and J. The DO utilization also increased steadily as the series of runs progressed up through Run H, which occurred in the peak of the algae season; and then decreased in Runs I through K as the lake water algae content decreased in the fall season.

Results of Direct, In-Line Filtration, Phase I

Since the emphasis of this research was on small treatment systems, the primary goal was to evaluate the simplest systems for high-quality surface waters. For that reason, the goal was to use only a single coagulant, either alum or a cationic polymer. In some filter runs using alum as a coagulant, pH was lowered to about 6.8 with sulfuric acid in hopes of achieving better results. The acid was needed because of the relatively high alkalinity of the quarry water (150-200 mg/L as $CaCO_3$), which buffered the pH above 7.5 even after alum addition. Most upland waters of low alkalinity would have the pH reduced sufficiently by the alum alone so that this added complexity would not be necessary. Otherwise, a cationic polymer could be used as a sole coagulant.

Also, in view of the small system emphasis, the range of filtration rates studied was limited to 6.6 to 16.1 m/h (2.7 to 6.6 gpm/sq ft). Higher rates were considered inappropriate for small systems.

Rapid mixing of the chemicals with the filter influent water was achieved by static mixers. No flocculation time was provided, but some detention after rapid mixing did exist in the influent hoses and in the water above the filter media. Because of the clarity of the raw water and the low doses of chemicals used, no visible floc particles were evident in the water above the filter media. Nevertheless, the evidence of destabilization was dramatized by the quality of the filtrate and the abrupt loss of quality if the chemical feed was terminated either intentionally or accidentally.

TABLE 19. DROP IN DISSOLVED OXYGEN CONCENTRATION ACROSS SLOW SAND FILTER (INLET DO--OUTLET DO)

Run	First Value (ΔDO, mg/L)	Mean of All Other Values (ΔDO, mg/L)	(Number)*	Last Value (ΔDO, mg/L)
A	1.5	0.78	(5)	0.7
B	1.0	0.97	(17)	2.8
C	2.2	2.6	(1)	4.3
D	3.5			6.3
E	2.5			5.4
F		4.0	(1)	
G	2.7			3.8
H		6.8	(1)	
I	6.5	5.3	(1)	4.1
J	2.9	2.5	(1)	3.0
K		2.7	(1)	

*Number of individual values used to calculate the mean value.

The experimental results are summarized in Table 20 for all observation runs which were not disturbed by mechanical problems or abrupt changes in the raw water quality which required mid-run corrections to the chemical feed level.

Numerous additional filter runs were made between the runs shown in Table 20. These were made to select proper chemical dosages prior to an observation run. All runs shown in Table 20 were operated with optimum chemical dosage, at least the best dosage using the methods of dosage selection discussed in the Procedures section of this report.

Influent and effluent turbidity data for the entire Phase I and II period of CRF operation are presented in Tables 20 and 21. Table 21 presents the average raw and average filtrate turbidity for each run up to the time of breakthrough. Breakthrough was defined arbitrarily as that time when the recorded effluent turbidity began to have a consistent increasing value. Table 21 includes the high turbidity at the beginning of the run, the average

TABLE 20. SUMMARY OF RAPID CONSTANT-RATE FILTER RUNS IN PHASE I

Run Number	Dates	Rate m/h	Avg. Raw NTU	Chemicals (mg/L)			Acid Used	pH Fil-trate	Avg. Fil-trate NTU	Break-through*		At End	
				Alum	Polymer	Cl_2				HL (cm)	Hrs.	HL (cm)	Hrs.
Alum Runs													
	(1981)												
A-1	10/20-22	6.8	5.2	7.1	---	---	yes	6.8	0.18	150	40	200	48
A-2	10/27-30	7.1	3.5	6.5	---	---	yes	6.3	0.19	none		195	76
A-3	11/3-5	7.1	3.2	6.4	---	---	yes	6.7	0.14	none		207	49
A-4	11/10-11	11.0	7.0	6.9	---	---	yes	6.8	0.21	none		208	31
A-5	11/17-18	11.5	3.7	7.0	---	---	yes	6.9	0.17	none		146	24
B-1	12/1-2	11.0	7.9	7.6	---	---	yes	6.8	0.28	none		186	30
Cat-Floc T Runs													
B-2	12/8-12	6.8	5.2	---	0.76	---	no	8.5	0.15	none		193	97
B-3	12/14-15	11.5	5.2	---	0.70	---	no	8.6	0.27	none		207	26
	(1982)				Lake frozen over beginning 12/18/81								
B-4	1/4-6	6.8	2.9	---	0.77	---	no	8.6	0.21	none		197	52
B-7	2/8-15	11.2	0.35	---	0.09	---	no	8.3	0.13	none		204	168
B-8	2/15-22	11.2	0.5	---	0.09	---	no	8.3	0.13	none		208	168

Snow melt runoff begins--lake open around edges, 2/21/82

TABLE 20. CONTINUED

Run Number	Dates (1982)	Rate m/h	Avg. Raw NTU	Chemicals (mg/L)			Acid Used	pH Fil-trate	Avg. Fil-trate NTU	Break-through*		At End	
				Alum	Polymer	Cl_2				HL (cm)	Hrs.	HL (cm)	Hrs.
Alum Runs													
B-10	2/24	11.2	4.5[†]	10.3 & 12.3	---	---	yes	6.8	0.33	136	6.5	---	---
B-11	3/1-2	6.6	5.1	10.8	---	---	yes	6.8	0.23	135	22	154	26
Cat-Floc T													
B-13	3/8-13	7.1	2.0	---	0.84	---	no	8.2	0.32	none		210	120
Lake ice cover completely gone, 3/28/82													
Alum Runs													
B-15	3/29-30	7.3	4.2	6.7	---	---	yes	6.8	0.24	none		130	26
C-1	4/13-14	7.3	3.6	8.1	---	---	yes	6.8	0.21	none		207	31
C-3	4/20-21	11.7	5.7	7.8	---	---	yes	7.6	0.35	130	5	210	13
C-4a	4/26-27	10.2	7.6	12.1	---	---	yes	6.5	0.48	130	7	170	11
Head loss recorders installed													
Alum Runs													
E-1	6/1-2	7.3	6.6	7.4	---	---	yes	6.8	1.2	65	4	158	22
Cat-Floc T-1 Runs													
G-1	6/23-25	6.8	3.7	---	0.80	---	no	8.5	0.55	none		108	47

TABLE 20. CONTINUED

Run Number	Dates (1982)	Rate m/h	Avg. Raw NTU	Chemicals (mg/L)			Acid Used	pH Fil-trate	Avg. Fil-trate NTU	Break-through*		At End	
				Alum	Polymer	Cl_2				HL (cm)	Hrs.	HL (cm)	Hrs.
Cat-Floc T Runs													
G-2	6/28-30	6.6	3.0	---	0.67	---	no	8.4	0.51	none		56	49
Alum Plus Cl_2 (Cl_2 in some runs)													
H-1	7/20-21	7.3	14.7	19.8	---	5	no	7.9	1.03	160	12	210	16
H-2d	7/27-28	7.3	18.1	9.4	---	no	yes	7.0	1.68	68	8	210	19
H-2f	7/30-31	7.3	13.3	12	---	5	yes	6.6	0.72	190	17	210	19
H-2h	8/1-2	8.3	12.5	10.3	---	5	yes	7.9	4.64	none		140	24
I-2	8/10-11	7.3	2.3†	8.3	---	5	yes	7.8	0.33	none		110	24
I-3	8/11-12	7.3	2.2*	8.3	---	3	yes	7.8	0.31	none		119	24
I-4	8/18	16.1	1.4	5.8	---	3	yes	7.9	0.23	none		210	11
I-5	8/19-20	16.1	1.2	5.6	---	3	yes	7.9	0.50	none		114	24
Cat-Floc T Runs													
I-6c	8/24-25	14.4	2.7	---	1.49	no	no	8.4	0.55	none		133	22
I-6e	8/27-29	12.4	1.9	---	1.21	no	no	8.4	0.45	none		155	45
Alum Runs													
J-1	9/2	13.2	8.2	8.8	---	---	yes	7.8	0.27	187	10	206	11
J-2	9/2-3	13.5	2.9	8.1	---	---	yes	7.8	0.28	none		200	10

TABLE 20. CONTINUED

Run Number	Dates (1982)	Rate m/h	Avg. Raw NTU	Chemicals (mg/L)			Acid Used	pH Fil-trate	Avg. Fil-trate NTU	Break-through*		At End	
				Alum	Polymer	Cl_2				HL (cm)	Hrs.	HL (cm)	Hrs.
Cat-Floc T													
J-3	9/9-10	13.3	2.3	---	0.48	---	no	8.5	0.34	none		151	28
J-4	9/10-11	13.4	2.0	---	0.53 (3 hr) 0.35 (31 hr)	---	no	8.5	0.34	none		158	34
Alum Runs													
J-6	9/15-16	7.7	3.2	6.1	---	---	yes	7.8	0.21	none		136	24
J-7	9/16-17	7.6	3.4	6.1	---	---	yes	7.8	0.20	none		142	27
Cat-Floc T													
J-8	9/22-26	7.7	2.5	---	0.58	---	no	8.4	0.32	none		201	95
J-9	9/27-10/2	7.7	1.7	---	0.45	---	no	8.4	0.29	none		161	122

*Breakthrough was defined as the time when effluent turbidity began to have a consistent increasing value.

†Avg. of high and low Ratio Turbidimeter readings only.

TABLE 21. TURBIDITY DATA FOR RAPID CONSTANT-RATE FILTER RUNS IN PHASES I AND II

			Raw Turbidity (NTU)			Filtrate Turbidity (NTU)		
Run Number	Water Temp. °C	Filtrate pH	High*	Avg.†	Low‡	High*	Avg.†	Low‡
Alum Runs at 6-8 m/h								
pH Controlled								
A-1	14	6.8	6.6	5.2	4.7	0.90	0.18	0.15
A-2	13	6.3	4.2	3.5	2.8	1.88	0.19	0.11
A-3	13	6.7	6.0	3.2	2.4	1.00	0.14	0.10
B-11	2	6.8	7.6	5.1	4.0	1.74	0.23	0.17
B-15	4	6.8	6.5	4.2	3.3	2.06	0.24	0.17
C-1	6	6.8	4.0	3.6	3.2	3.00	0.21	0.16
E-1	17	6.8	6.7	6.5	6.5	2.50	1.19	0.86
H-2d	28	7.0	20.0	18.1	16.1	16.0	1.68	1.10
				$\bar{x}$ = 6.18			0.51	(92%)[§]
pH Uncontrolled								
J-6	20	7.8	4.4	3.2	2.4	0.44	0.21	0.18
J-7	20	7.8	4.8	3.3	2.2	0.44	0.20	0.17
				$\bar{x}$ = 3.25			2.05	(94%)
With Cl_2 and pH Controlled								
H-2f	27	6.6	14.5	13.3	10.0	3.10	0.72	0.31
				$\bar{x}$ = 13.3			0.72	(95%)
With Cl_2 and pH Uncontrolled								
H-1	27	7.9	20.2	14.7	11.3	8.50	1.03	0.39
H-2h	27	7.9	14.0	12.5	10.8	6.50	4.64	2.40
I-2	25	7.8	2.6[¶]	---	1.9[¶]	0.94	0.33	0.26
I-3	25	7.8	2.4[¶]	---	2.0[¶]	0.91	0.31	0.22
				$\bar{x}$ = 7.9			1.58	(80%)
Alum Runs at 11-16 m/h								
pH Controlled								
A-4	13	6.8	7.8	7.0	6.1	1.50	0.21	0.14
A-5	12	6.9	4.4	3.7	3.3	1.41	0.17	0.09
B-1	7	6.8	11.1	7.9	6.2	2.05	0.28	0.11

TABLE 21. CONTINUED

Run Number	Water Temp. °C	Filtrate pH	Raw Turbidity (NTU)			Filtrate Turbidity (NTU)		
			High*	Avg.†	Low‡	High*	Avg.†	Low‡
B-10	2	6.8	$4^{\P}$	---	$5^{\P}$	1.05	0.33	0.19
C-4a	10	6.5	8.6	7.6	6.9	2.10	0.48	0.40
				$\bar{x}$ = 6.55			0.29 (96%)	
pH Uncontrolled								
C-3	7	7.6	7.0	5.7	5.2	1.60	0.35	0.22
J-1	24	7.8	16.0	8.2	1.9	0.81	0.27	0.20
J-2	24	7.8	6.4	2.9	2.0	0.79	0.28	0.22
				$\bar{x}$ = 5.60			0.30 (95%)	
With Cl_2 and pH Uncontrolled								
I-4	25	7.9	2.1	1.4	1.0	0.66	0.23	0.13
I-5	25	7.9	1.4	1.2	0.9	0.74	0.50	0.40
				$\bar{x}$ = 1.3			0.37 (72%)	
Cat-Floc Runs at 6-8 m/h								
B-2	5	8.5	9.6	5.2	3.7	0.93	0.15	0.10
B-4	4	8.6	9.1	2.9	2.0	1.05	0.21	0.16
B-13	3	8.2	3.7	2.0	1.6	1.52	0.32	0.24
G-1	21	8.5	6.6	3.7	1.2	1.05	0.55	0.38
G-2	23	8.4	9.5	3.0	1.4	1.04	0.51	0.42
J-8	18	8.4	3.9	2.5	1.8	1.22	0.32	0.25
J-9	17	8.4	4.6	1.7	0.9	0.73	0.29	0.21
				$\bar{x}$ = 3.00			0.34 (89%)	
Cat-Floc Runs at 11-16 m/h								
B-3	4	8.6	5.8	5.2	4.5	2.46	0.27	0.19
B-7	3	8.3	1.0	0.35	0.3	0.60	0.13	0.09
B-8	3	8.3	1.8	0.50	0.3	0.88	0.13	0.12
I-6c	24	8.4	6.0	2.7	1.6	1.28	0.55	0.38

TABLE 21. CONTINUED

Run Number	Water Temp. °C	Fil-trate pH	Raw Turbidity (NTU) High*	Avg.[†]	Low[‡]	Filtrate Turbidity (NTU) High*	Avg.[†]	Low[‡]
I-6e	22	8.4	2.3	1.9	1.2	0.92	0.45	0.35
J-3	23	8.5	3.7	2.3	1.6	0.76	0.34	0.27
J-4	23	8.5	2.3	1.9	1.3	0.83	0.34	0.27
				$\bar{x}$ = 2.12			0.32 (85%)	

[*] Highest value at beginning of filter run.

[†] Avg. for entire run up to time of breakthrough.

[‡] Lowest value of run.

[§] Percent reduction in the average value for each group of runs.

[¶] No continuous turbidity record. Values based on grab samples measured on Ratio Turbidimeter.

for the remainder of the run up to the time of breakthrough, and the lowest turbidity at any time during the run, which usually occurred near the end of the run. The data in Table 21 are presented in chronological sequence but grouped according to common chemical pretreatment and filtration rate. Water temperature data are added to emphasize the seasonal change occurring over the series of runs.

To round out the data presentation, the particle count results in the 7-12 μm size range and total coliform results are summarized in Tables 22 and 23 and the run length data in Table 24 for the Phase I and II CRF study period.

In addition to the tables, the data from typical runs will be presented in a series of figures, along with a discussion of seasonal trends and comparisons.

Results, Autumn 1981--

In the fall of 1981, the quarry water was of reasonably good quality with low algal populations as evidenced by low chlorophyll measurements. Either alum with pH adjusted to 6.8, or Cat-Floc T were quite adequate to achieve good filtrate turbidity. Typical results are shown in Figs. 37 through 40 for one alum run and one Cat-Floc T run.

TABLE 22. MEAN PARTICLE REMOVAL BY RAPID FILTER IN 7-12 μm SIZE RANGE

				Mean % Particle Removal			
				First Hour		Remainder	
Season	Run Dates	Chemical Used	Mean Influent No./mL	%	(#)*	%	(#)*
	1981						
Fall	10/20-12/15	Alum	2320	97.6	(5)	98.8	(5)
		Cat-Floc T	1170	91.9	(2)	96.7	(2)
	1982						
Winter (Ice Covered)	1/4-2/22	Cat-Floc T	370	68.7	(3)	87.0	(3)
Snow Melt	2/24-3/13	Alum	2190	97.0	(1)	99.0	(1)
		Cat-Floc T	1620	97.0	(1)	98.0	(1)
Spring (Ice Gone)	3/29-4/21	Alum	2860	92.0	(3)	94.0	(3)
Summer	6/1-8/18	Alum	13040	85.0	(1)	99.0	(1)
		Cat-Floc T & T-1	1350	89.0	(2)	85.5	(2)
		Alum & Cl_2	2730	86.0	(3)	92.0	(3)
Fall	9/2-10/2	Alum	1640	94.0	(2)	96.5	(2)
		Cat-Floc T	340	87.0	(2)	87.5	(2)

---

*Number of mean filter run values used to calculate the mean % removal value.

TABLE 23. MEAN TOTAL COLIFORM REMOVAL BY RAPID FILTER

Season	Run Dates	Chemical Used	Mean Influent No./100 mL	Percent Coliform Removal			
				First Hour		Remainder	
				%	(#)*	%	(#)*
	1981						
Fall	10/20-12/15	Alum	1300	90.5	(4)	91	(3)
		Cat-Floc T	8200	88	(2)	96.5	(2)
	1982						
Winter (Ice Covered)	1/4-2/22	Cat-Floc T	1500	77.7	(3)	89.7	(3)
Snow Melt	2/24-3/13	Alum	1600	93	(1)	96	(1)
		Cat-Floc T	640	72	(1)	89	(1)
Spring (Ice Gone)	3/29-4/21	Alum	350	79	(3)	91.3	(3)
Summer	6/1-6/30	Alum	90	80	(1)	86	(1)
		Cat-Floc T & T-1	50	81.5	(2)	86	(2)
Fall	9/2-10/2	Alum	550	86.5	(2)	89	(2)
		Cat-Floc T	170	70.5	(2)	86.5	(2)

*Number of mean filter run values used to calculate the mean percent removal value.

TABLE 24. MEAN RUN LENGTHS FOR RAPID FILTER COMPARING CAT-FLOC T AND ALUM

Season	Run Dates	Chemical	Mean Run Length (h)	
			At 7.3 m/h (3 gpm/ft^2)*	At 12.2 m/h (5 gpm/ft^2)*
Fall '81	10/20-12/15	Alum	54	28
		Cat-Floc T	95	26
Winter '81-'82 (Ice Covered)	1/4-2/22	Cat-Floc T	52	168†
Snow Melt	2/24-3/13	Alum	22	6.5
		Cat-Floc T	120	no data
Spring	3/29-4/21	Alum	29	6
Summer	6/1-8/18	Alum	4	no data
		Cat-Floc T & T1	48	no data
		Alum + Cl_2	21	17
Fall	9/2-10/2	Alum	26	10
		Cat-Floc T	109 "	31

*Nominal rates, actual rates somewhat higher or lower as shown in Table 20.

†Mid-winter with extremely good raw water (Runs B-7 & B-8).

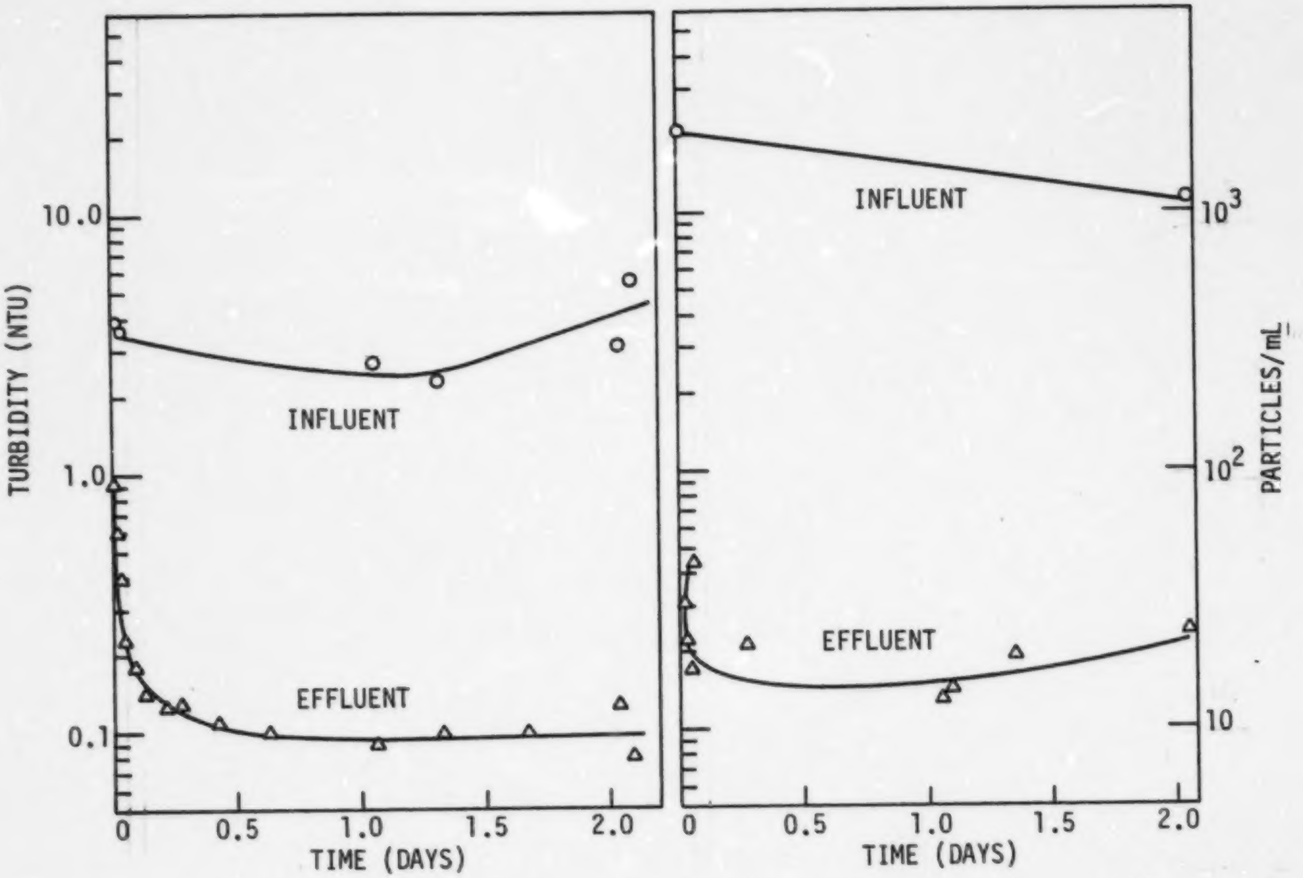


Fig. 37. Turbidity and 6.9-12.5 μm particle count during Run A-3 at 7.1 m/h (2.9 gpm/ft^2) using alum coagulant.

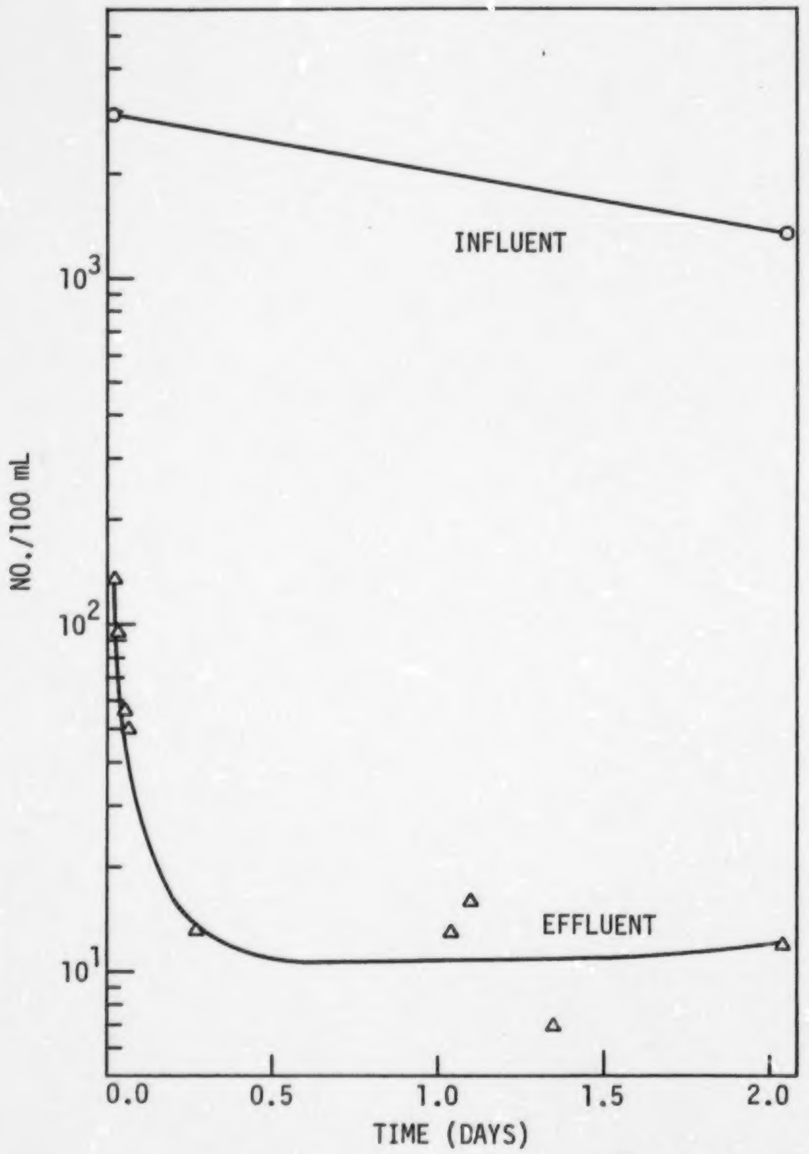


Fig. 38. Total coliform bacteria during Run A-3 at 7.1 m/h ($2.9\ gpm/ft^2$) using alum coagulant.

7.1 m/h

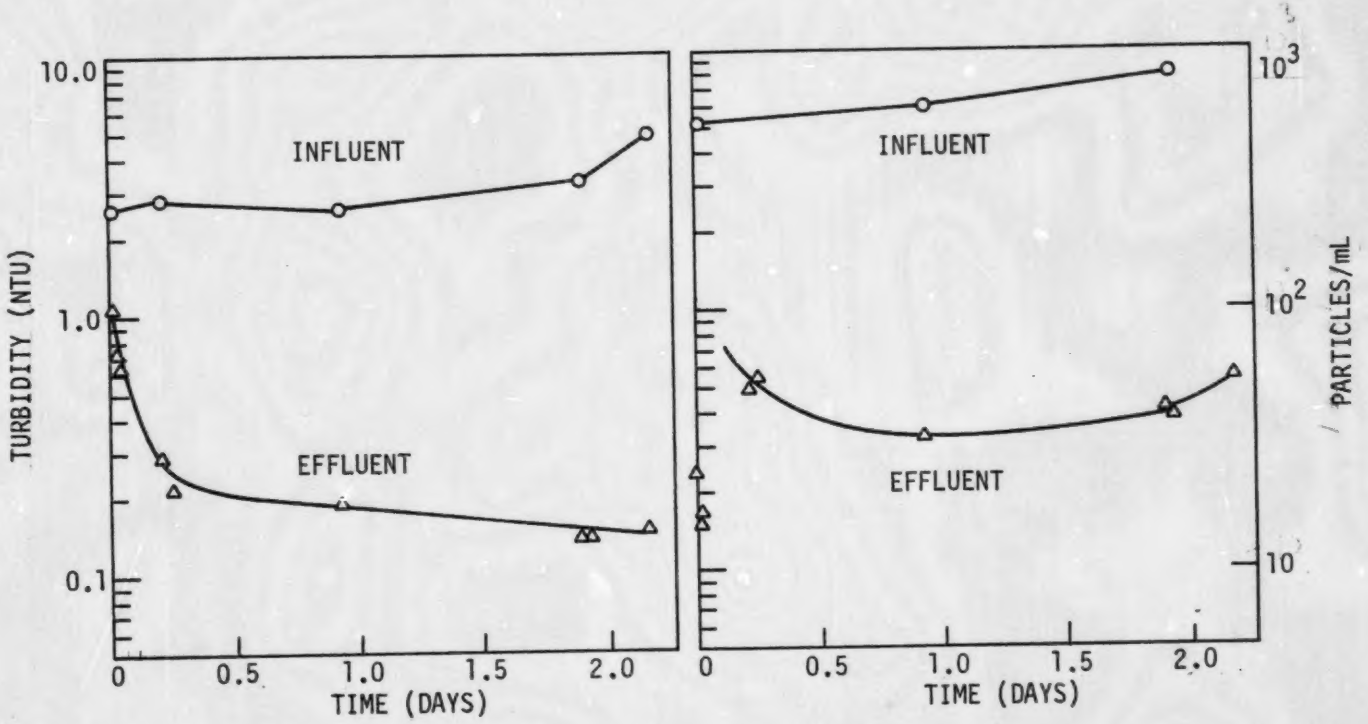


Fig. 39. Turbidity and 6.9-12.5 µm particle count during Run B-4 at 6.8 m/h (2.8 gpm/ft^2) using Cat-Floc T coagulant.

31

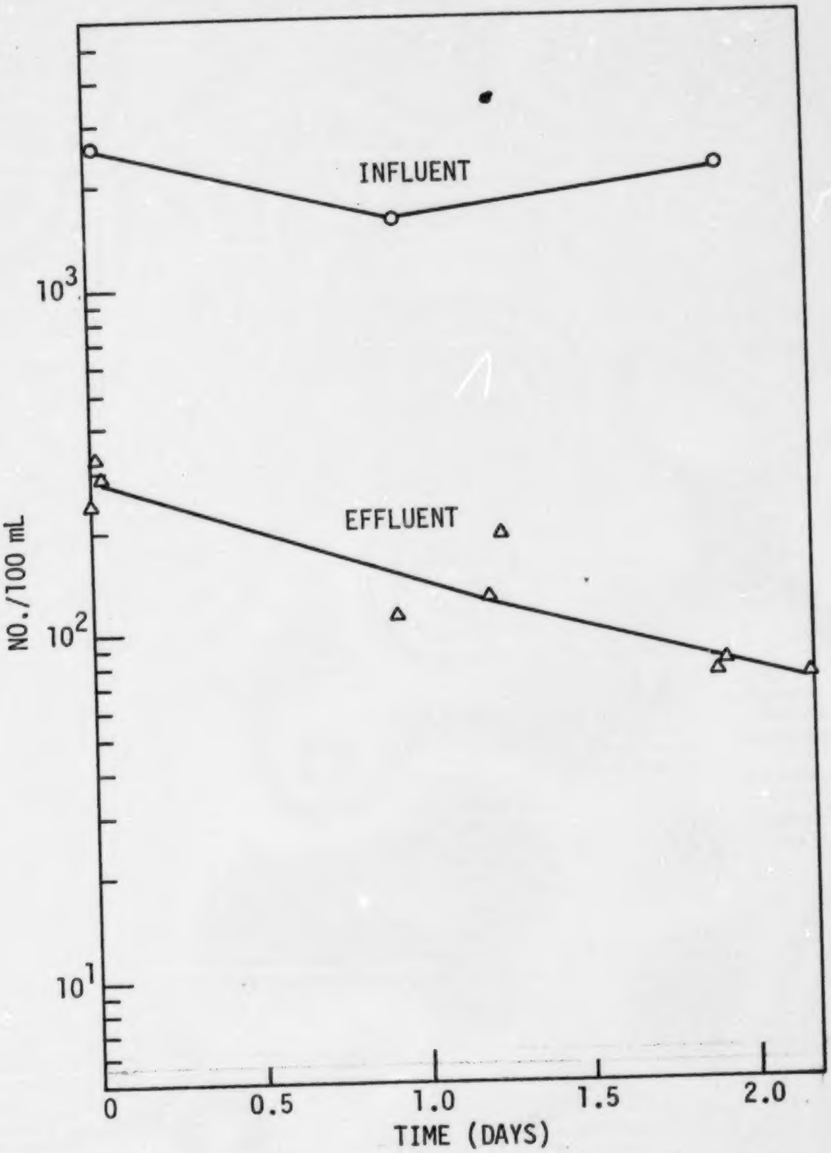


Fig. 40. Total coliform bacteria during Run B-4 at 6.8 m/h (2.8 gpm/ft^2) using Cat-Floc T coagulant.

The initial improvement period was pronounced in all runs, and most clearly defined by the continuous turbidity recording. It is clear that several hours are required for the filtrate to approach a steady quality.

Results, Winter 1981-82--

With the formation of the ice cover on the quarry, the raw water quality got progressively better. Only Cat-Floc T was used during the winter months; this is regrettable, in hindsight. Long runs with good filtrate were obtained with extremely low dosage of the polymer (0.09 mg/L). One typical run (B-7) is shown in Figs. 41 and 42.

With this high quality raw water, the percentage removal of any of the three parameters was not as good as with poorer raw water, but the absolute levels of turbidity and particle count were excellent in spite of the lower fractional removals.

Comparing the winter and snow melt periods of Table 23, it is evident that alum appears superior to Cat-Floc T in total coliform removal, a trend also evident in Table 22 for particle removal efficiency.

Results, Spring 1982--

With the onset of snow melt runoff into the quarry, but with ice cover still prevailing, the raw water immediately became more difficult to treat. Whereas average filtrate turbidities of 0.15 to 0.20 were commonly achieved in the fall and winter (Table 20), it was not possible to achieve such results during this period of partial ice cover. Higher alum dosages were used in an attempt to improve the filtrate (as in Runs B-10 and B-11) but this resulted in terminal breakthrough of turbidity with short filter cycles.

The use of Cat-Floc T during this period generally eliminated the terminal breakthrough problem, but the average filtrate turbidity was 0.32 NTU as shown in Fig. 43 and Table 20 for Run B-13.

After the ice had completely left the quarry, the turbidity results with alum were different in shape with a shorter initial improvement period; the average filtrate turbidity of 0.21 NTU as shown in Fig. 44 for Run C-1 was somewhat better than obtained with Cat-Floc T in the prior Run B-13.

Results, Spring and Summer 1982--

The first major algae bloom occurred in late April and resulted in short cycles to breakthrough as shown in Fig. 44 for Run C-3.

These difficult treatment conditions persisted to varying degrees throughout the summer with the worst runs observed in late May, early June, and late July, as shown in Table 20. In Run E-1 while using alum, the average filtrate turbidity was 1.2 NTU and terminal breakthrough began in 4 hours at 65 cm head loss. Two different Cat-Floc polymers were used in Runs G-1 and G-2, and an average filtrate turbidity of 0.55 and 0.51 NTU was achieved, respectively, but run length was better at two days.

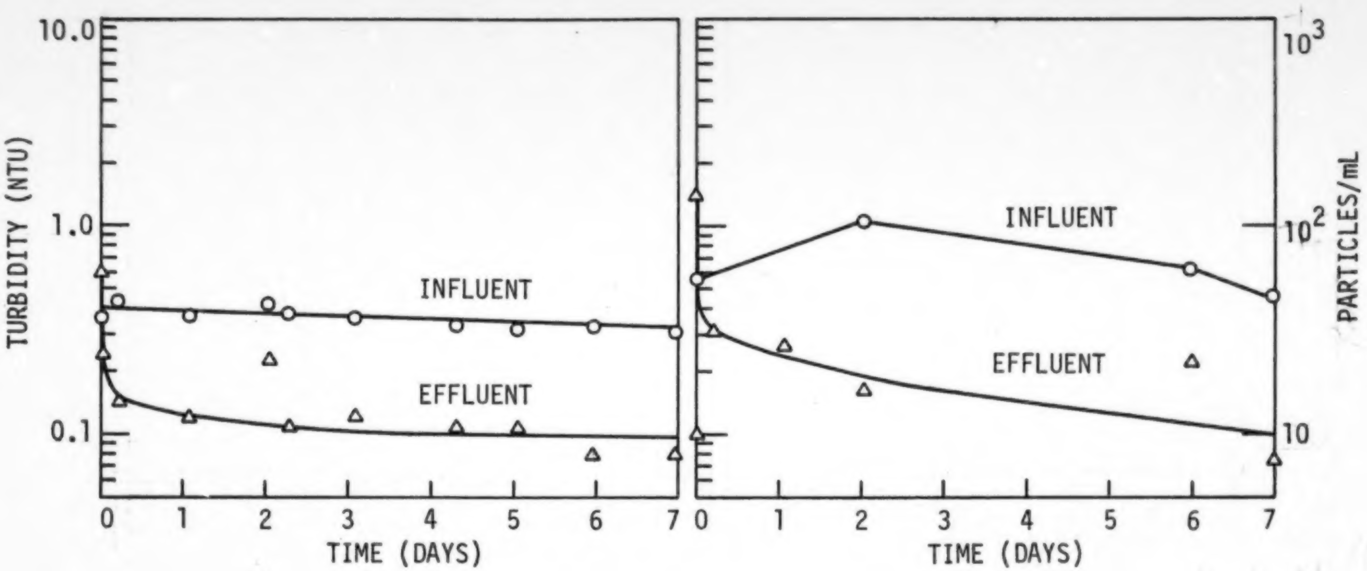


Fig. 41. Turbidity and 7-12 μm particle count during Run B-7 at 11.2 m/h (4.6 gpm/ft2) using Cat-Floc T coagulant under winter ice at 2 °C.

41

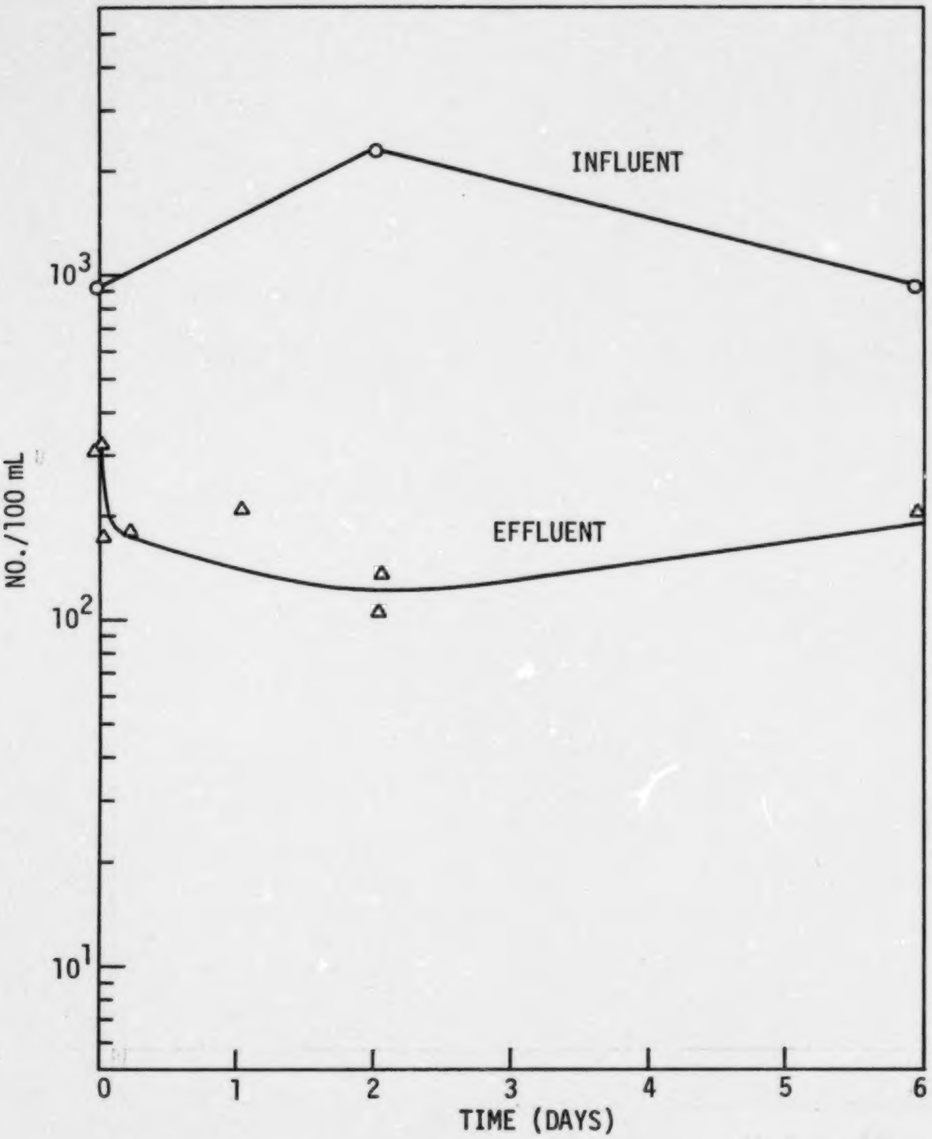
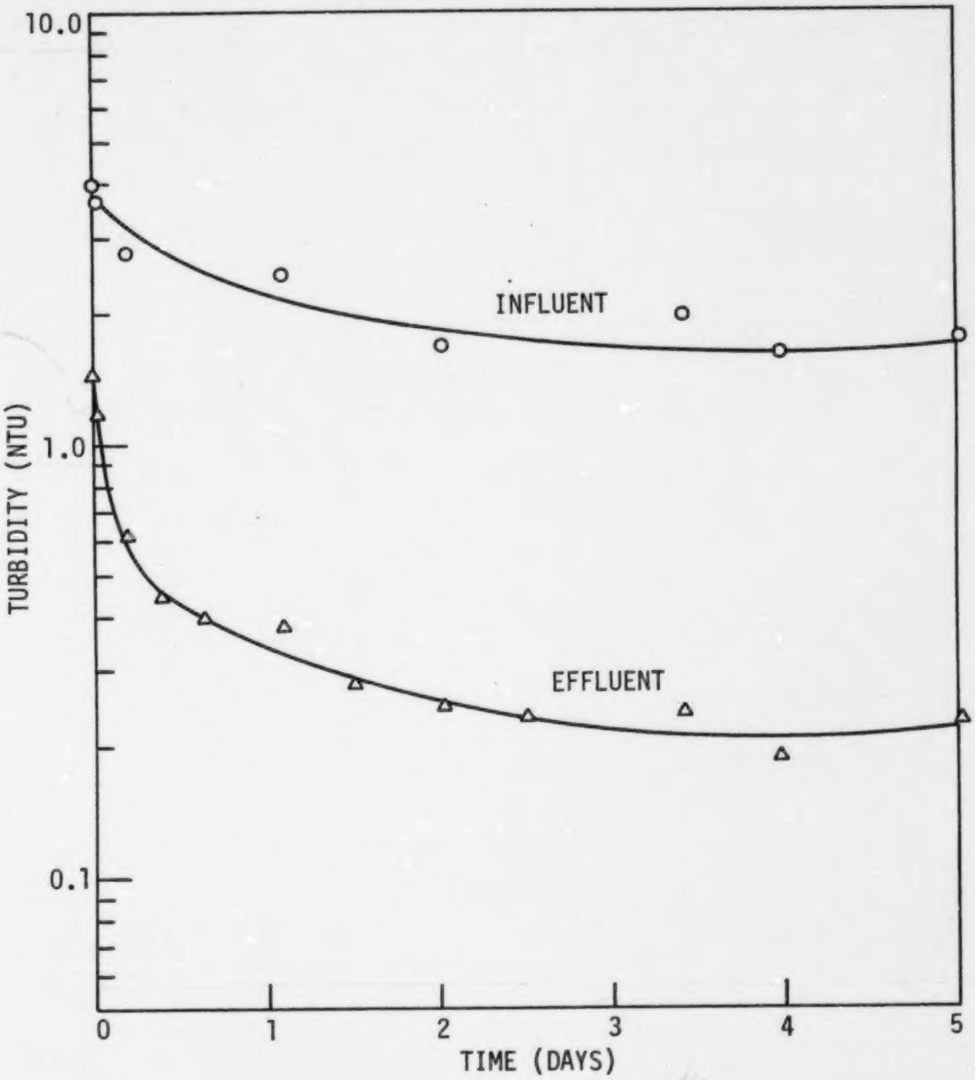


Fig. 42. Total coliform bacteria during Run B-7 at 11.2 m/h (4.6 gpm/ft^2) using Cat-Floc T coagulant.



7.1 m/h

Fig. 43. Turbidity during Run B-13 at 7.1 m/h (2.9 gpm/ft^2) using Cat-Floc T coagulant.

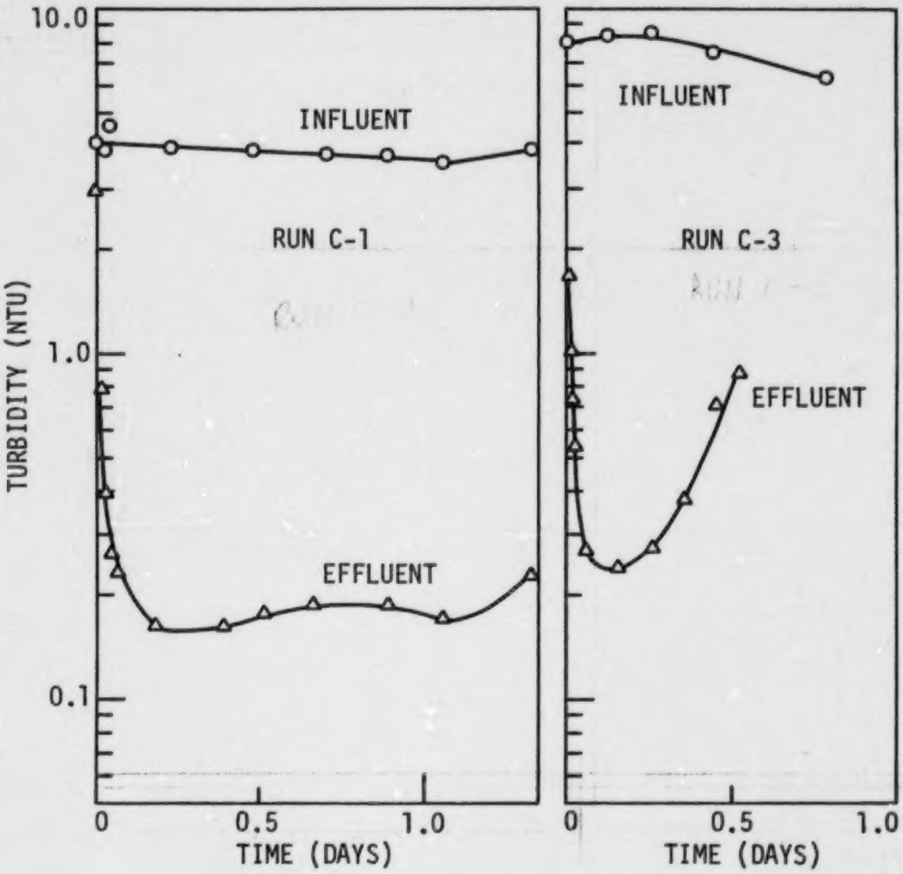


Fig. 44. Turbidity during Run C-1 and C-3 using alum coagulant. Run C-1 at 7.3 m/h (3 gpm/ft^2), Run C-3 at 11.7 m/h (4.8 gpm/ft^2).

In late July during a severe algal bloom, it became impossible to produce acceptable filtrate without the use of chlorine ahead of the filters as is evident by comparing Runs H-1 and H-2f with Cl_2 against Run H-2d without Cl_2. These runs using alum plus Cl_2 were only marginally acceptable because of short run lengths. Run H-2h did not produce acceptable average turbidity even with the use of Cl_2 and a fairly high alum dosage of 10.3 mg/L. Of course, prechlorination or preozonation are common practices in direct filtration plants. The use of prechlorination was avoided in this research because of the desire to use bacterial parameters of removal efficiency. No bacterial data collection was attempted during the periods of prechlorination.

Results, Autumn 1982--

After July, the quarry water improved dramatically, achieving the lowest raw water turbidities of the year, except during winter ice cover. In spite of the apparent good raw water during August and September of 1982, it was impossible to achieve filtrate turbidity levels as low as in the fall of 1981. During the J series runs in September 1982, low average turbidities of 0.20 to 0.34 were achieved, with alum being superior to Cat-Floc T. The same trend is evident in the particle count data and total coliform data (Tables 22 and 23, respectively).

RESULTS OF PHASE II, CRF VS DRF FILTRATION

Filter Run Summary

The comparison of constant-rate filtration vs declining-rate filtration actually began on June 11, 1982. However, a number of problems developed in the first two months of operation which caused rejection of the DRF data. These included: (i) inadequate backwash outlet capacity, corrected by enlarging the outlet connections; (ii) neglecting to skim the sand on installation of the filter media, corrected by replacing all the dual media in the DRFs and CRF including proper skimming procedures; (iii) failure to achieve equal head loss development under identical operating conditions, corrected by adopting a regular bed compaction procedure; and (iv) difficulty due to poor raw water and rapidly changing raw water during the summer, corrected by allowing the lake to stabilize by the end of August 1982.

Subsequent to correction of these problems, eight valid filter runs were completed between Sept. 2 and Oct. 2, 1982, which are presented and discussed in this chapter.

There were two runs each at nominal rates of 7.70 m/h (3.15 gpm/ft^2) and 13.35 m/h (5.46 gpm/ft^2) with alum and with Cat-Floc T addition. A summary of these eight runs is shown in Table 25. In most cases, the chemical dosage was optimized in the preliminary run. A correction in polymer dosage was made in the initial hours of runs J-4 and J-9. Sulfuric acid was added in the alum runs to reduce the influent pH to a level which, it was hoped, would yield a more favorable filtration condition [3].

TABLE 25. SUMMARY OF FILTER RUN PARAMETERS IN PHASE II

Run Number	Dates (1982)	Average Flow Rate (m/h)	Average Raw Water Turbidity (NTU)	Chemical Dosage (mg/L)	Water pH	
					Raw	Filter Effluent
J-1 (A)[*]	9/2	13.2	8.17	8.84 alum + 3 mL/L acid[†]	8.4	7.8
J-2 (A)[*]	9/2-3	13.5	2.89	8.12 alum + 3 mL/L acid[†]	8.5	7.8[‡]
J-3 (P)[*]	9/9-10	13.3	2.32	0.48 polymer	8.5	8.5
J-4 (P)[*]	9/10-12	13.4	1.95	2.75 hr-0.53 polymer 31 hr-0.35 polymer	8.5	8.5
J-6 (A)[*]	9/15-16	7.7	3.17	6.06 alum + 4 mL/L acid[†]	8.4	7.8
J-7 (A)[*]	9/16-18	7.6	3.35	6.11 alum + 4 mL/L acid[†]	8.4	7.8
J-8 (P)[*]	9/22-26	7.7	2.54	0.58 polymer	8.4	8.4
J-9 (P)[*]	9/27-10/2	7.7	1.66	2.0 hr-0.56 polymer 120 hr-0.45 polymer	8.4	8.4

[*] (A) = alum (dosage expressed as standard ground filter alum), (P) = Cat-Floc T.

[†] Concentrated H_2SO_4 in stock tank of alum stock solution.

[‡] No actual data; estimated value.

Sample Run Data

Run J-3 will be used to illustrate the data collected during each run. Figures 45 and 46 show the influent and effluent turbidities during Run J-3. Figure 46 includes the turbidity of the combined effluent of all four DRF filters and the individual effluent of DRF #3. Note that the turbidity of DRF #3 is very nearly the same as the CRF both in magnitude and shape. The downward trend during the run was typical of all polymer runs and terminal turbidity breakthrough was not observed. The influent turbidity increased about mid-run. This increased the CRF effluent slightly but had a greater effect on the DRF because DRF #1 was backwashed shortly after this increase in turbidity. The peak DRF effluent turbidity as a result of the backwash of DRF #1 was 0.53 NTU for the combined effluent as compared to an average of 0.42 NTU following the other 3 backwashes. Similar influent/effluent turbidity relationship occurred in Runs J-1 and J-7.

Particle count data for Run J-3 are shown in Fig. 47. These are graphs of the 7-12 μm sized particles only. As explained earlier, the actual particle count data included all particles from 1 to 60 μm in twelve channels. Run J-3 was about 28 hours long and only 2 influent samples were taken for particle analysis. Therefore, the variations in influent quality that appear on the turbidity graph are not evident on the particle count graphs. The initial sample on the CRF effluent was probably taken before the backwash water had been flushed from the filter; therefore, the count is lower than the counts of the remainder of the run. The DRF graph reveals the relationship of one DRF filter (#3) effluent to the combined DRF effluent. DRF #3 was backwashed at the beginning of the run and the first 4 sets of effluent samples were taken before DRF #4 was backwashed. The last 3 sample sets in the run were taken just prior to BW #1, BW #2 and BW #3, respectively and indicate the result of the increased influent turbidity, which is also shown in Fig. 46. Both the CRF and the DRF demonstrated about a 1-log reduction in particle count.

Figures 48 and 49 present the bacterial results obtained for Run J-3. These results cannot be correlated to the influent turbidity or particle count values because the influent was 'spiked' with sewage and therefore was independent of the quarry influent water quality. Attempts were made to maintain high influent bacterial counts by using primary and secondary wastewater effluent from the Ames Wastewater Treatment Plant. Secondary effluent less than two days old seemed to work best, but even this source did not give very high coliform counts consistently.

Figure 48 suggests about a 1-log reduction in total coliform count for both the CRF and DRF. The average influent count was 165/100 mL. The average CRF and Combined DRF effluents were 27/100 mL and 21/100 mL, respectively. The DRF #3 and CRF effluents follow the decreasing trend similar to the turbidity performance curves.

Figure 49 demonstrates the unusual standard plate count data for the runs. The influent numbers follow a decreasing trend as did the influent total coliform numbers. However, the effluent data for all filters were very erratic, especially when the effluent counts exceeded the influent counts.

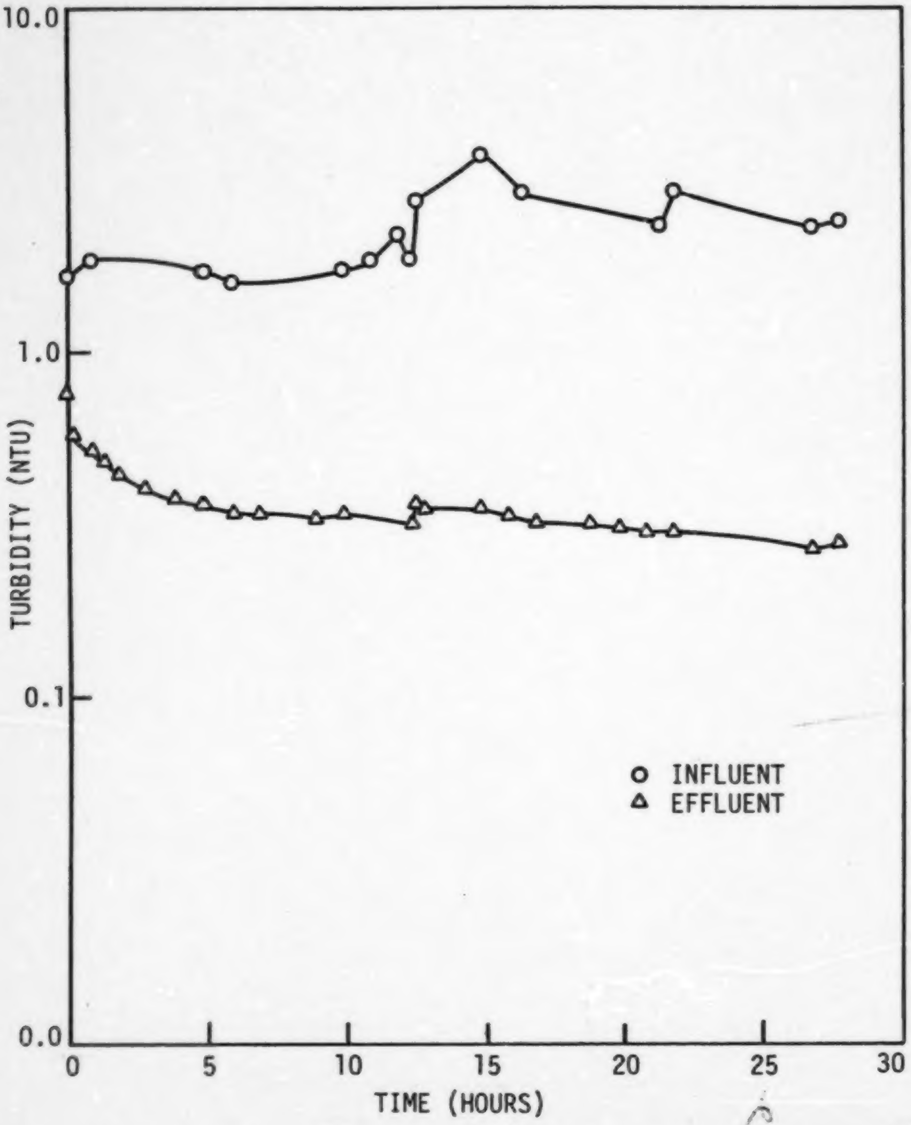


Fig. 45. Turbidity for CRF during Run J-3 at 13.3 m/h (5.34 gpm/ft^2) using Cat-Floc T coagulant.

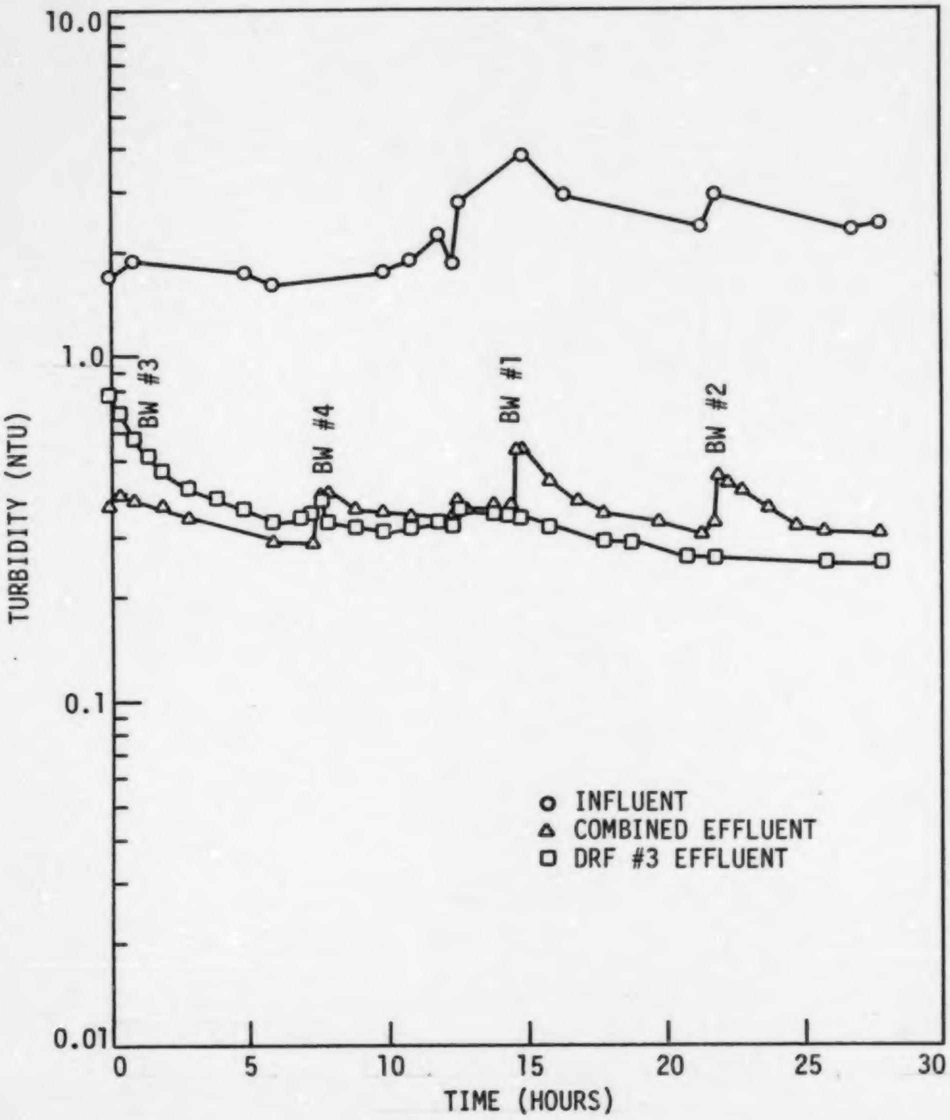


Fig. 46. Turbidity for DRF during Run J-3 at 13.3 m/h (5.34 gpm/ft^2) using Cat-Floc T coagulant.

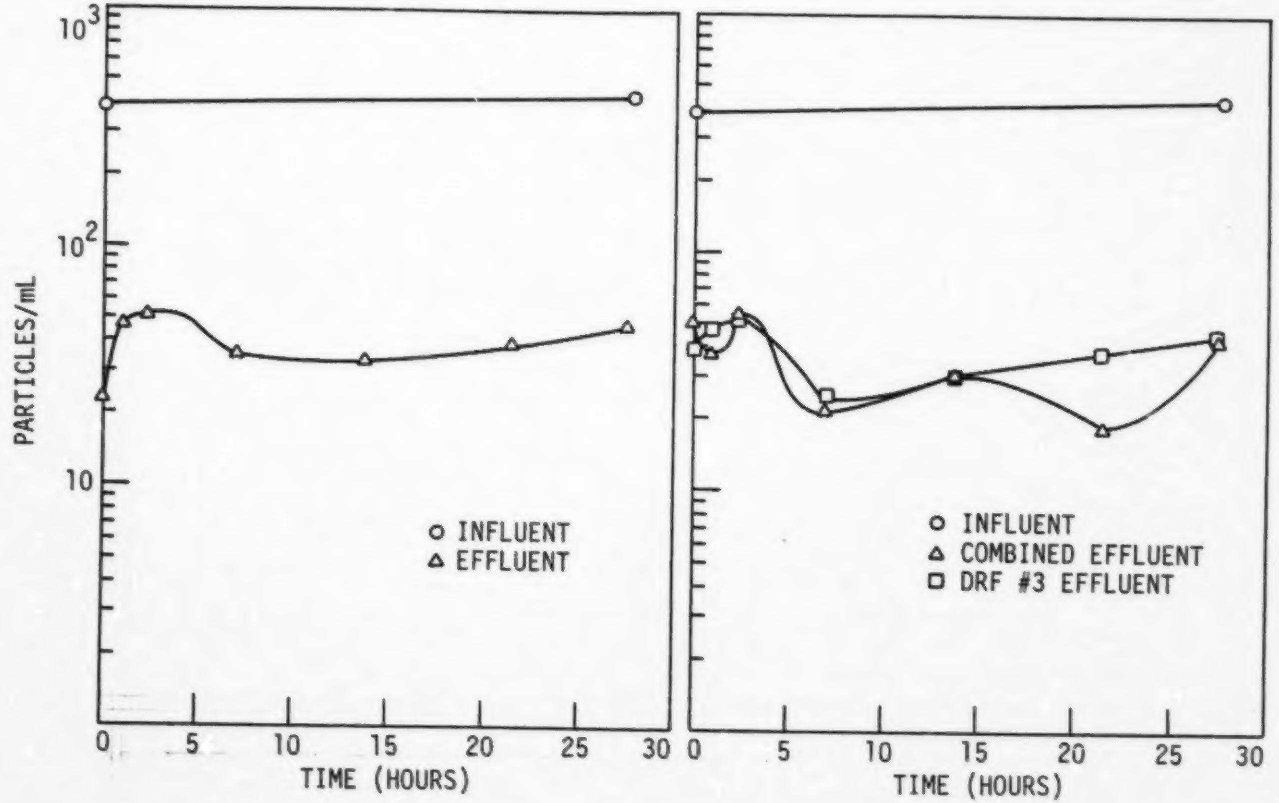


Fig. 47. 7- to 12-μm particle count for CRF (left graph) and DRF (right graph) during Run J-3 at 13.3 m/h (5.34 gpm/ft^2) using Cat-Floc-T coagulant.

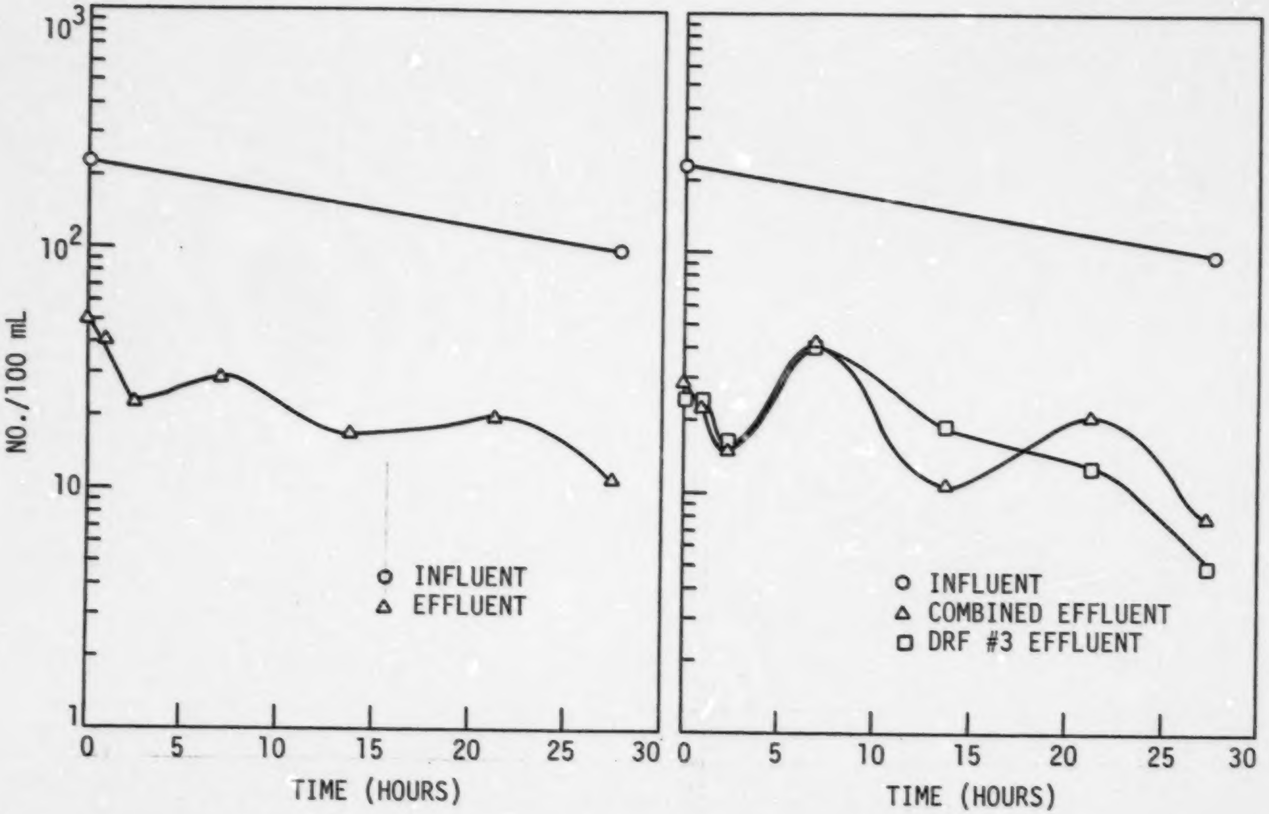


Fig. 48. Total coliform count for CRF (left graph) and DRF (right graph) during Run J-3 at 13.3 m/h (5.34 gpm/ft^2) using Cat-Floc T coagulant.

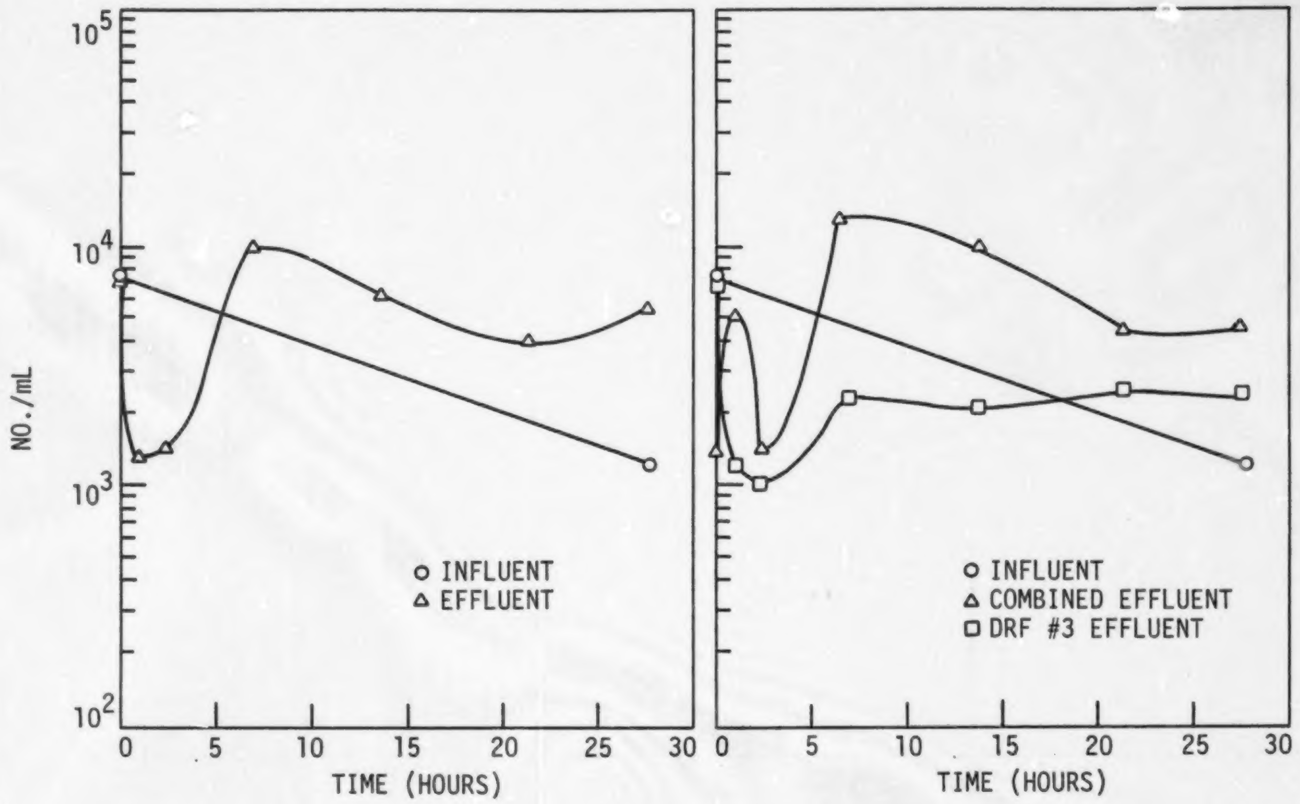


Fig. 49. Standard plate count for CRF (left graph) and DRF (right graph) during Run J-3 at 13.3 m/h (5.34 gpm/ft^2) using Cat-Floc T coagulant.

It was theorized that there was bacterial growth in the media and effluent piping. All clear plastic effluent piping was exchanged for stainless steel tubing early in the project. Also, periodic runs were made with chlorine addition to disinfect the system. Neither of these efforts significantly changed the plate count results as is seen in this later project data.

Figure 50 shows the head loss data for Run J-3 along with the DRF mean values of the four operating levels for the run (levels as designated in reference 18). During the CRF initial improvement period, the head loss increase per unit time was less than for the remainder of the run. This was not evident in the DRF profile since it reflects the combined condition of the 4 DRFs in various stages of clogging. The effect is more noticeable on polymer Runs J-3 and J-4 (Figs. 50 and B-3 [Appendix B]) than on alum runs J-1 and J-2 (Figs. B-1 and B-2, Appendix B).

On some occasions, the 4 individual DRF flows were recorded before, during and 50 to 60 minutes after a backwash of one of the filters. Figure 51 represents the flow vs time relationship for all 4 DRF during the backwash of DRF #3 in Run J-9. The 3 filters remaining in service during the backwash show increasing flow (common increasing water level) during the backwash period and then decreasing flow to their new and lower flow rates after the backwashed filter was put back into operation. In Run J-9, DRF #3 was put back into service at level 4 at about 1.46 times the average flow rate. The effluent piping turbulent head loss had been adjusted prior to the series of runs at an average filtration rate of 7.70 m/h (3.15 gpm/ft^2) with a goal of limiting the maximum rate at level 3 to 1.5 times the average flow rate.

The remainder of the turbidity and head loss graphs comprise Appendices B and C, respectively. The 7-12 µm particle count and total coliform count graphs for Run J-9 are in Appendix C. Three additional graphs of individual DRF flow during typical backwash are given in Appendix D.

Summary of Filtration Run Results

Introduction--

The data from the eight runs were summarized into a series of tables from which key parameter comparisons were made. These key parameters were turbidity, particle count, flow and head loss, total coliform count, standard plate count, and chlorophyll-a. As explained in the Procedures section, the coagulant dosage was optimized prior to any formal runs. Corrections to the coagulant dosage were made during formal Runs J-4 and J-9.

Turbidity--

Table 26 gives the turbidity results. Influent and effluent turbidities were recorded continuously. Therefore, by taking all of the high, low, and inflection points from the recorder paper and correcting them to actual turbidity values via correlation to the periodic Ratio Turbidimeter readings, a set of continuous turbidity values was generated. These data were then reduced to average turbidity values for the entire run by numerical analysis. If breakthrough occurred, the average values are for the total run up to the

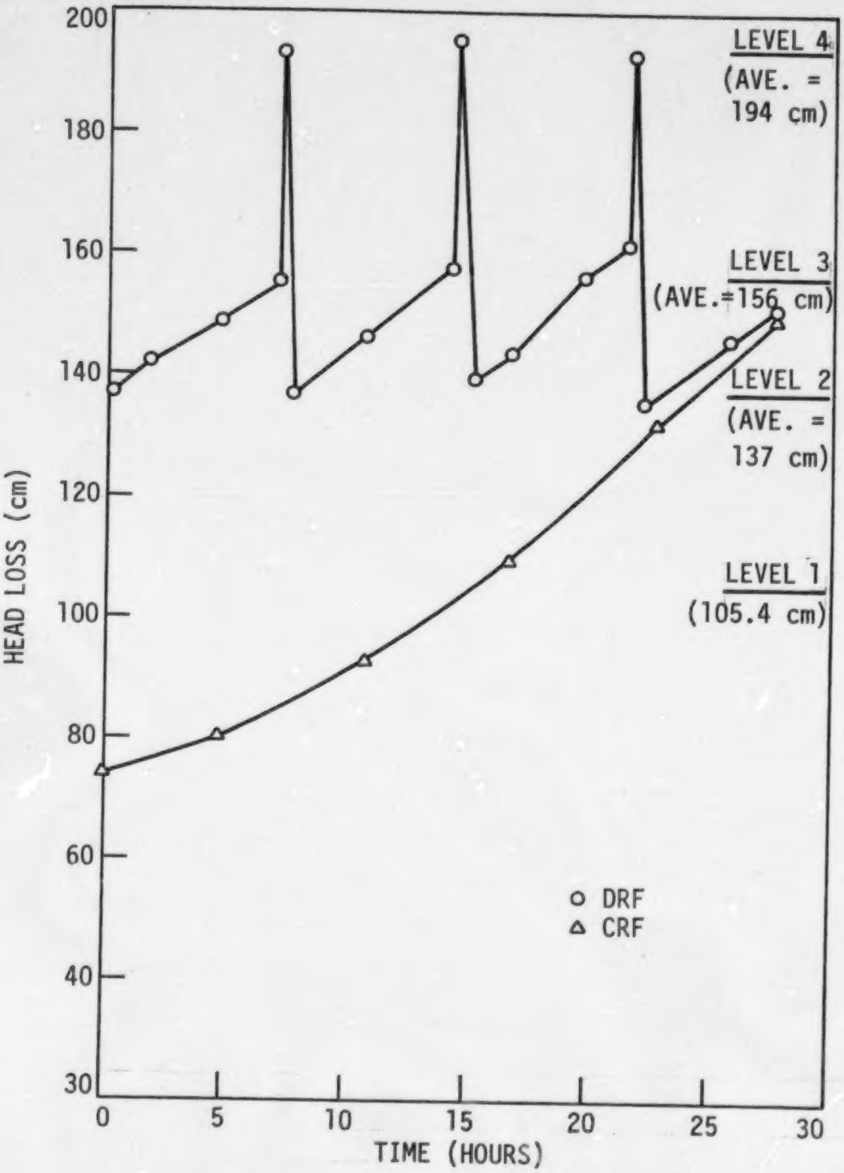


Fig. 50. Head loss for CRF and DRF during Run J-3 at 13.3 m/h (5.34 gpm/ft^2) using Cat-Floc T coagulant. Levels 1, 2, 3 and 4 are DRF designations from Cleasby and DiBernardo [18] (feet = cm × 0.0328).

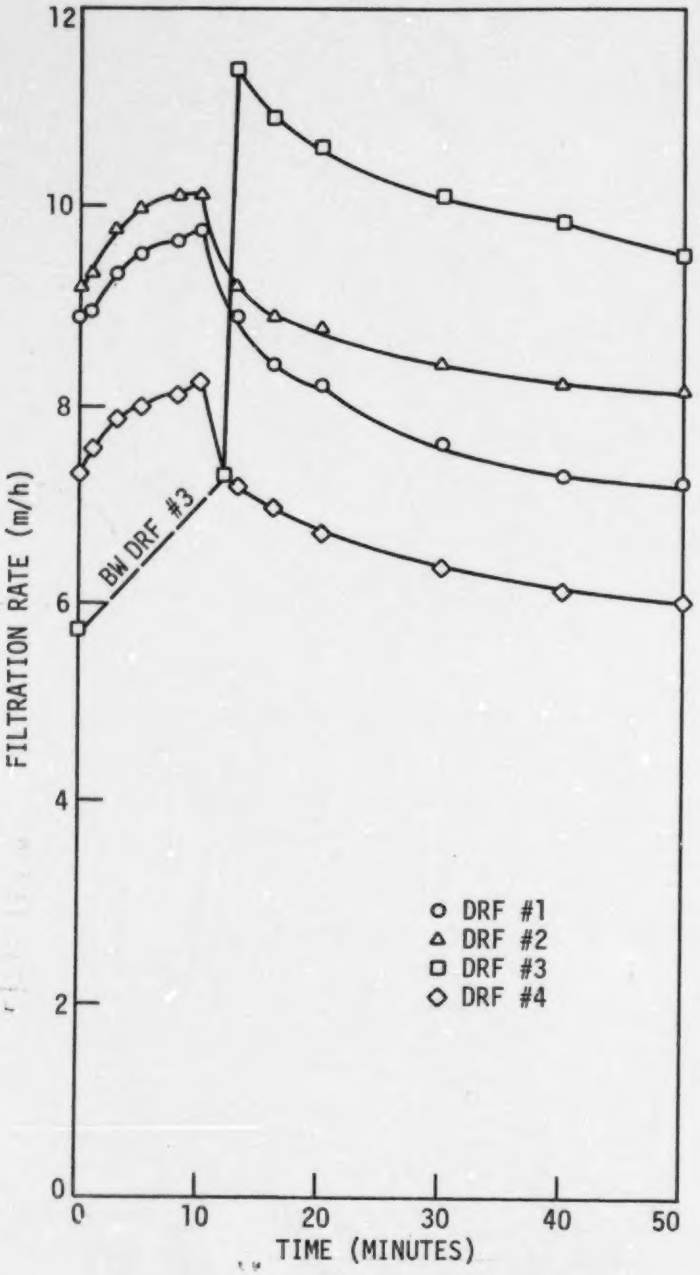


Fig. 51. Changes in filtration rates for the four DRF during the backwash of DRF #3 on Oct. 2,1982 (during Run J-9). Backwash begins at time = 0. Mean flow before backwash was 7.78 m/h (3.12 gpm/ft^2) and mean flow at time = 60 minutes had returned to 7.75 m/h (3.11 gpm/ft^2) (gpm/ft^2 = m/h x 0.402).

TABLE 26. SUMMARY OF FILTER RUN TURBIDITIES

Run Number	Run Length (h)	Influent Turbidity (NTU)			CRF Turbidity (NTU)			DRF Turbidity (NTU)*		
		High	Average	Low	High	Average	Low	High	Average	Low
Nominal 13.3 m/h (5.44 gpm/ft^2)										
J-1 (A)[†]	11.5	16.0	8.2	1.9	0.81	0.27	0.20	0.86	0.29	0.19
J-2 (A)	10.0	6.4	2.9	2.0	0.79	0.28	0.22	0.49	0.28	0.21
J-3 (P)	27.8	3.7	2.3	1.6	0.76	0.34	0.27	0.53	0.35	0.29
J-4 (P)	34.0	2.3	1.9	1.3	0.83	0.34	0.27	0.43	0.34	0.27
Nominal 7.70 m/h (3.15 gpm/ft^2)										
J-6 (A)	24.0	4.4	3.2	2.4	0.69	0.21	0.17	0.44	0.21	0.18
J-7 (A)	26.9	4.8	3.3	2.2	0.61	0.20	0.16	0.44	0.20	0.17
J-8 (P)	95.2	3.9	2.5	1.8	1.22	0.32	0.25	0.72	0.33	0.26
J-9 (P)	121.7	4.6	1.7	0.9	0.73	0.29	0.21	0.54	0.29	0.21

[*] For the combined effluent of the DRF bank of four filters.

[†] (A) = alum, (P) = Cat-Floc T.

time of breakthrough. Run J-1 had the highest average influent turbidity. This average value was probably low because there were several periods when the indicator and recorder were off scale.

From Table 26 it can be seen that there were notable differences in alum and polymer run average and low effluent turbidity values. Independent of the influent turbidity, the average effluent turbidity for the alum runs was about 0.08 NTU lower than the average effluent turbidity for the polymer runs. From the data given in Table 26, there is no apparent difference in CRF and DRF effluent turbidities.

Figures 45 and 46 presented previously and Figs. A-1 through A-14 in Appendix A were included for a full turbidity record of the 8 runs. The alum runs show about 1 to 1.1 log reduction in turbidity and the polymer runs show about 0.75 to 0.80 log reduction. A 1-log reduction is equal to 90% turbidity removal.

Figures A-1 and A-2 in Appendix A show the filter responses to a significant increase in influent turbidity. The CRF, which was well established in the run when the increase occurred, gave a moderately increasing effluent turbidity followed by breakthrough at the end of the run. DRF #1 and #2 which were backwashed prior to the increase in influent turbidity probably would have effluent turbidity profiles similar to the CRF. However, DRF #3, which was backwashed shortly after the increase in influent turbidity, shows the effect of the higher initial filtration rate after backwash upon the filtrate quality. The post backwash effluent turbidity of DRF #3 starts high and does not recover to an effluent turbidity equal to the CRF effluent turbidity until near the end of the run. The Combined DRF effluent increased along with the DRF #3 effluent and thereafter maintained a higher average turbidity than during the first two segments of the run prior to the increase in influent turbidity.

Particle count--

Table 27 is a summary of the total and 7-12 μm particle count data for the 8 runs. The samples taken during the first hour of the run were tabulated separately since there was usually a dramatic improvement during this period. In Runs J-1 and J-6, the first-hour values for DRF #3 were in the middle of the full run consisting of 1 backwash of each of the 4 filters because DRF #3 was backwashed third and second in the 4 filter sequence, respectively (see Figs. A-2 and A-8 in Appendix A).

Particle count data summarized in Table 26 are similar to much of the turbidity data. For Run J-1, the CRF values are lower than the DRF #3 values or the Combined DRF values because of the mid-run influent turbidity increase. The CRF and Combined DRF particle count values are very similar in Runs J-6 and J-9. The turbidity graphs for Run J-3 (Figs. 45 and 46) do not reinforce the particle count data, which suggests that the Combined DRF effluent is better.

Figure 47 shows an influent-to-effluent particle count reduction of about 1 log cycle. On the DRF portion of Fig. 47, the DRF #3 particle count

TABLE 27. PARTICLE COUNT SUMMARY* (values in No./mL)

	Total Particle Count		7-12 μm Particle Count	
Run Number	First Hour	Remainder of the Run	First Hour	Remainder of the Run
J-1				
Influent:	21,772 (1)		2484 (1)	
Effluent:				
CRF	3311 (2)	2438 (5)	59 (2)	70 (5)
DRF	2297 (2)	2996 (5)	41 (2)	92 (5)
#3 DRF	4911 (1)	2351 (6)	124 (1)	82 (7)
J-3				
Influent:	14,403 (2)		408 (2)	
Effluent:				
CRF	3459 (2)	3356 (4)	35 (2)	41 (4)
DRF	2477 (2)	2890 (5)	44 (2)	34 (5)
#3 DRF	3775 (2)	3052 (5)	43 (2)	39 (5)
J-6				
Influent:	42,919 (2)		803 (2)	
Effluent:				
CRF	4439 (2)	1665 (4)	70 (2)	41 (4)
DRF	4064 (2)	1664 (4)	59 (2)	42 (4)
#3 DRF	5896 (2)	1928 (5)	69 (2)	50 (5)
J-9				
Influent:	14,715 (2)		270 (2)	
Effluent:				
CRF	3095 (2)	2040 (6)	47 (2)	41 (6)
DRF	1944 (2)	1907 (6)	41 (2)	37 (6)
#3 DRF	3523 (2)	1855 (6)	54 (2)	40 (6)

---

*Number in parentheses following each particle count is the number of samples averaged in the particle count.

data points do not decrease during the run as the DRF #3 flow decreases. Figure C-1 (Appendix C) does show the decreasing particle counts which follow the decreasing DRF #3 flow in Run J-9. However, the influent particle count also decreased during the run.

Flow and head loss--

Flow data were collected on the DRF and CRF to ensure equal areal flows to both filter types. These data are given in Table 28. A numerical analysis was conducted on the flow rate vs time data recorded during each run. Although the flow rates were essentially constant, the flow variations were taken into account in calculating the average values. The average difference in DPF and CRF average run flow rates as a percent of the lower flow rate was 0.9%.

TABLE 28. AVERAGE FLOW RATE AND HEAD LOSS INCREASE VALUES

	Flow Rate (m/h)		Head Loss Increase (cm/h)	
Run Number	CRF	DRF	CRF	DRF*
J-1 (A)†	13.18	13.20	11.6	11.6
J-2 (A)	13.45	13.50	12.0	12.4
J-3 (P)	13.30	13.33	2.8	3.1
J-4 (P)	13.57	13.25	2.5	2.8
J-6 (A)	7.73	7.70	4.3	4.1
J-7 (A)	7.65	7.56	4.0	4.1
J-8 (P)	7.68	7.78	1.8	1.7
J-9 (P)	7.70	7.78	1.0	1.1

*Additional head loss was induced in Runs J-6 through J-9 by adjusting the DRF control valves as discussed under Procedures.

†(A) = alum, (P) = Cat-Floc T.

Table 28 also gives the rate of head loss increase values for the CRF and DRF for each run. DRF media compaction was recognized as a factor in obtaining comparable head loss results so the backwash procedure was adjusted to ensure comparable compaction (as discussed previously under Procedures). The head loss increase values in Table 28 for the constant rate filter were determined by taking the total head loss increase from start of the run to run termination divided by the run length. For the DRF, the 4 'linear' portions of the head loss profiles between backwashes were averaged to determine the head loss increase values. From the data given in Table 28, there appears to be no apparent difference in the CRF and DRF head loss increases.

The declining-rate filter bank head loss curve consisted of 4 levels. These levels are indicated in Fig. 50 and summarized in Table 29 for all eight runs. These levels have been described by Cleasby and DiBernardo [18]. Level 1 is the water level when all filters are clean and operating at the average filtration rate. Level 2 is the lowest level just after backwash of 1 filter when all filters are operating in equilibrium. Level 3 is the highest 4-filter operating water level just prior to a filter backwash. Level 4 is the peak 3 filter water level during the backwash of the 4th filter.

As described in Procedures, the clean filter flow rate was restricted to 150% of 7.3 m/h (150% of 3.0 gpm/ft^2) at level 3 during the filter runs with a nominal rate of 7.3 m/h. The nominal flow rate for Runs J-6 to J-9 was 7.7 m/h (3.15 gpm/ft^2), which is slightly greater than the estimated mean flow rate assumed during the flow restriction calibration. The average clean filter flow rate at level 4 determined from Figs. 51 and D-1 through D-3 (Appendix D) was 1.60 × the average flow rate before the backwash. Also, the average filter flow rate for the cleanest filter just prior to backwash determined from the same figures was 1.31 times the average flow rate, which existed just before the backwash. Therefore, the level 3 clean filter flow rate was probably about 1.3 to 1.4 times the average flow rate, although it is not possible to determine the value from these figures.

Total coliform count and standard plate count--

Table 30 is a summary of the total coliform count data for Runs J-1, J-3, J-6 and J-9. The total coliform data are divided into first-hour and remainder-of-the-run averages as was done for the particle count data. The first-hour values for DRF #3 are from the first hour after the backwash of DRF #3. For Runs J-1 and J-6, the first-hour values are mid-run because DRF #3 was backwashed third and second during the full filter run, respectively. The full filter run covers the time span from the backwash of any one filter of the bank of filters until the next backwash of the same filter.

In Runs J-1, J-3 and J-9, the "first-hour" coliform counts are all higher than the "remainder-of-run" values for the same run. All of the "remainder-of-run" total coliform values for the CRF and DRF are very close for any particular run except for the DRF #3 value in Run J-9 which was somewhat higher than the CRF and DRF values. The effluent coliform counts in Run J-6 are all surprisingly high relative to those of the other runs. As with the other runs, the sewage used for spiking the influent was fresh about two hours prior to beginning the run. However, only one influent coliform

TABLE 29. DECLINING-RATE FILTER BANK WATER LEVELS (all values in cm and average of 4 readings unless noted)

Run Number*	BW Interval (h)	Level 1[†]	Level 2	Level 3	Level 4
J-1 (A)[‡]	2.88	101.5	157.5	183.5	211.8
J-2 (A)	2.50		156.5	181.5	212.0
J-3 (P)	6.94	105.4	137.0	156.0	194.0
J-4 (P)	8.50		137.0	158.5	198.0
J-6 (A)	6.00	86.4[§]	114.8	132.5	157.8
J-7 (A)	6.73		117.3	137.5	163.5
J-8 (P)	23.79	86.4	127.0	165.5	169.5
J-9 (P)	30.42		129.3	163.8	185.8

[*] J-1 thru J-4 at nominal rate of 13.3 m/hr, J-6 thru J-9 at nominal rate of 7.7 m/h.

[†] Only one value could be obtained in two sequential runs (level 1 at 7.7 m/h would have been 52.1 cm without effluent flow restriction).

[‡] (A) = alum, (P) = Cat-Floc T.

[§] Estimate.

TABLE 30. TOTAL COLIFORM COUNT SUMMARY*

Run Number	Influent (No/mL)	Effluent (No/mL)	
		First hour†	Remainder of Run
J-1 (A)‡	625 (2)		
CRF		55 (2)	15 (4)
DRF		18 (2)	14 (5)
DRF #3		42 (1)	17 (7)
J-3 (P)	165 (2)		
CRF		46 (2)	20 (5)
DRF		26 (2)	19 (5)
DRF #3		24 (2)	19 (5)
J-6 (A)	480 (1)		
CRF		98 (2)	96 (3)
DRF		82 (2)	91 (3)
DRF #3		106 (2)	91 (4)
J-9 (P)	170 (5)		
CRF		53 (2)	26 (5)
DRF		31 (2)	27 (5)
DRF #3		41 (2)	34 (5)

---

*Number in parentheses indicates the number of samples averaged in the total coliform count number.

†First-hour means first hour after BW. For DRF #3 this hour occurs mid run in Runs J-1 and J-6.

‡(A) = alum, (P) = Cat-Floc T.

sample was analyzed; therefore, the influent results may not have been representative.

Figures 48 and C-2 (Appendix C) give the total coliform count data for Runs J-3 and J-9, respectively. The effluent counts follow the influent count trends very well in all figures. In both runs, DRF #3 was backwashed at the beginning of the run and therefore had 4 stepwise decreasing flow rates throughout the run. The DRF #3 coliform counts show this decreasing trend very well in Fig. 48 but not as well in Fig. C-2.

Table 31 gives a condensed summary of the standard plate count data. In 7 of the 12 effluent values, the plate count exceeded the average influent plate count. The standard plate counts are very erratic as illustrated in Fig. 49 for Run J-3 and as shown in Table 31. The effluent counts were in many cases greater than the influent counts.

TABLE 31. STANDARD PLATE COUNT SUMMARY (values in No./mL)*

		Effluent		
Run Number	Influent	CRF	DRF #3	Combined DRF
J-1	40000 (2)	3500 (6)	1200 (8)	1400 (7)
J-3	4400 (2)	5000 (7)	2600 (7)	5700 (7)
J-6	3200 (1)	5800 (5)	4800 (6)	3000 (5)
J-9	3500 (5)	2200 (7)	1200 (7)	1400 (7)

*Number in parentheses indicates the number of samples averaged in the standard plate count value.

Chlorophyll-a--

Table 32 shows the chlorophyll-a removal in three of the runs. Only one set of chlorophyll samples was taken sometime in the middle of each run. There appeared to be little difference in CRF and DRF removal of chlorophyll-a with one exception in Run J-7.

Interpretation of Results

Initial Improvement Period--

The wasting of the initial water from a clean filter eliminates the poorer quality water during the initial improvement period. Water from

TABLE 32. CHLOROPHYLL-A IN FILTER INFLUENT AND EFFLUENT (mg/m^3) AND PERCENT REMOVAL

		Effluent					
		CRF		DRF #3		Comb. DRF	
Run Number	Influent	Conc.	% Rem.	Conc.	% Rem.	Conc.	% Rem.
J-3	2.9	1.2	59	1.2	59	1.2	59
J-7	3.8	0.7	82	0.8	79	1.6	58
J-9	4.2	1.5	64	1.6	62	1.7	60

filter to waste is directed either back into the plant for retreatment or to the sewer. The amount of water that is wasted must be balanced against the turbidity level at which the filter to waste is terminated and the overall benefit in effluent turbidity and bacterial reduction.

Table 33 shows the results of filter to waste calculations on the CRF and DRF #3 for three of the runs. Runs J-3, 4, and 9 were chosen because the CRF and DRF #3 were backwashed and put into operation at the beginning of the run.

From Table 33, it is apparent that the benefit from wasting a volume of water from the clean filter would result in only 0.01 to 0.02 NTU reduction in the average effluent turbidity. The period of filter-to-waste was from 1.08 to 2.8 hours, and as high as 7.2% of the entire run water volume was wasted. The filter to waste was defined arbitrarily to waste until the effluent dropped to 0.5 NTU during the initial improvement period of a newly backwashed filter. Both the time and the percent water wastage appear to be excessive when considering full-scale plant operation. The percent wastage was always higher for DRF #3 because it was being operated at about 1.25 to 1.47 times the average filter flow when steady state was established after the backwash. Figures 52 and 53 show the effect of wasting the first hour's filtered water on the total particle count in Run J-6. The first-hour 7-12 µm effluent particle counts (Fig. 52) are from 0.05 to 0.3 log cycles higher than those of the remainder-of-the-run (Fig. 53). The first-hour total particle count values are about 0.5 log cycle greater than those of the remainder-of-the-run values. DRF #3 shows the greatest change from first-hour to remainder-of-the-run values due to the high initial rate.

The choice of primary coagulant--

Alum and cationic polymer were used as primary coagulants during the eight filter runs. Table 26 shows that, independent of the influent

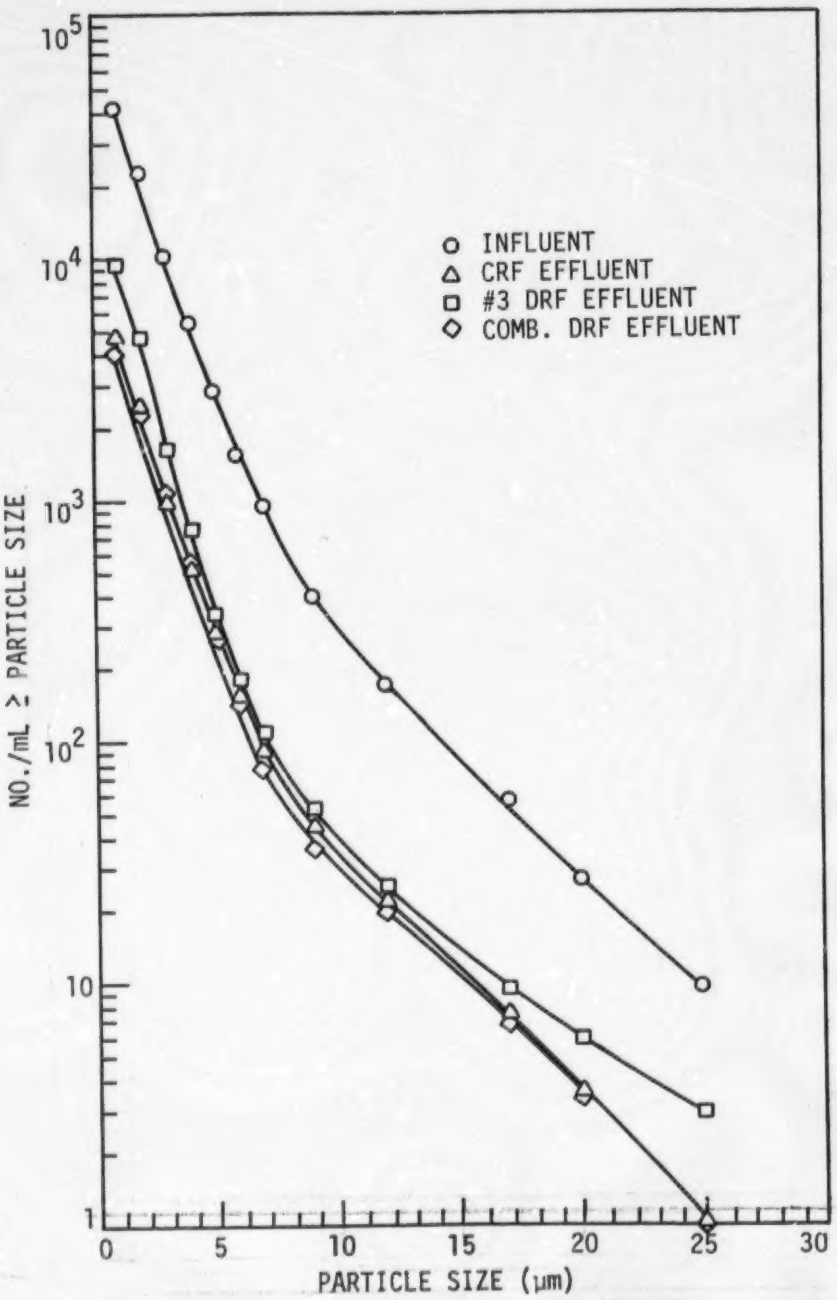


Fig. 52. Cumulative particle count for all 12 channels during the first hour of Run J-6 at 7.7 m/h (3.1 gpm/ft^2) using alum coagulant. The average of two samples is given for each point.

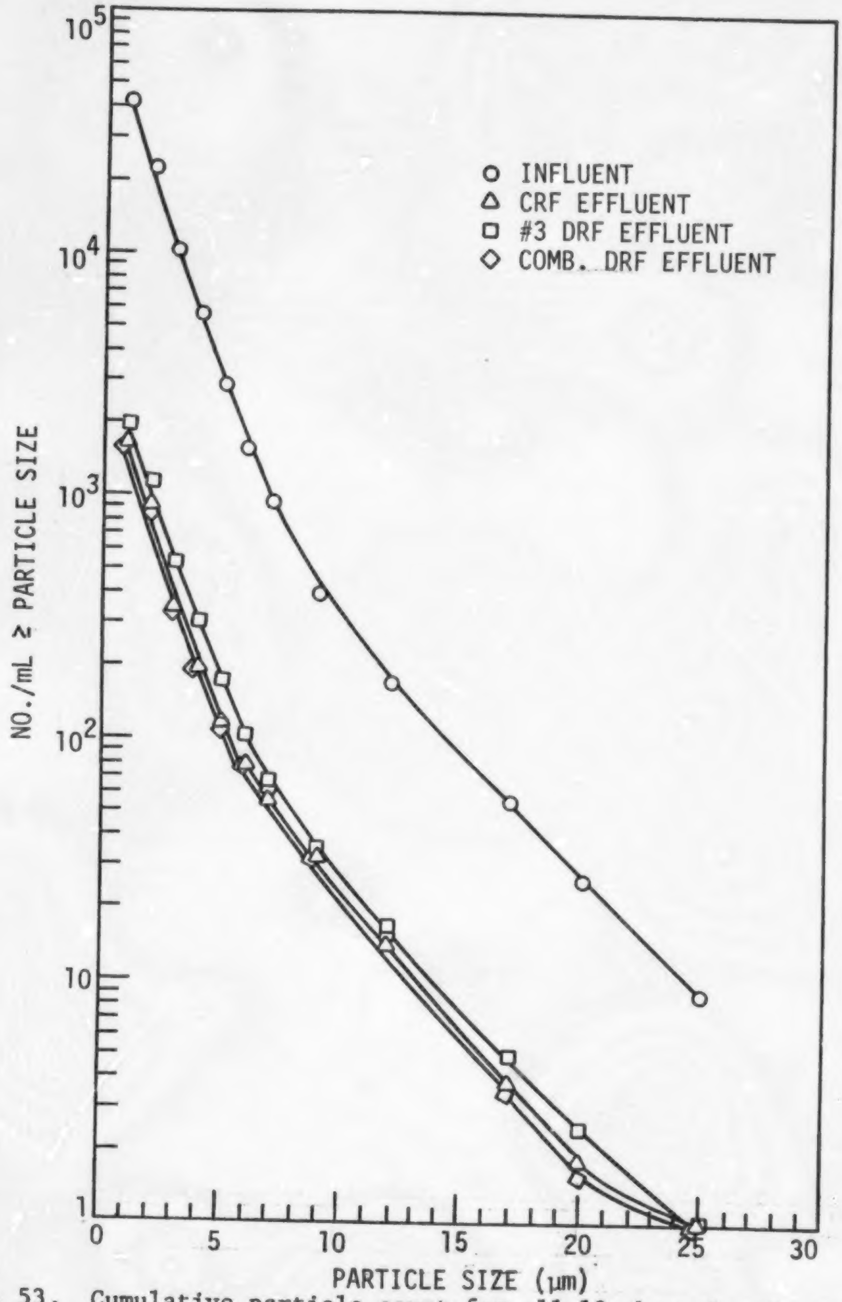


Fig. 53. Cumulative particle count for all 12 channels after the first hour of Run J-6 at 7.7 m/h (3.1 gpm/ft^2) using alum coagulant. The average of 2 influent, 4 CRF and Combined DRF and 5 DRF #3 samples is given for each point.

TABLE 33. FILTER TO WASTE (F.T.W.) SUMMARY (filter to waste was terminated when the effluent turbidity decreased to 0.5 NTU)

Run Number	Initial Impr. Period (h)	Avg. Effl. Turbidity w/o F.T.W.	Avg. Effl. Turbidity w/F.T.W.	% Water* Wasted
J-3				
CRF	1.08	0.34	0.33	3.9
DRF #3	1.45	0.33	0.32	7.2
J-4				
CRF	1.38	0.34	0.33	4.1
DRF #3	1.33	0.31	0.29	5.4
J-9				
CRF	2.25	0.29	0.28	1.8
DRF #3	2.80	0.27	0.26	3.2

$$^*\text{Water wasted} = \frac{\text{(hours wasted)(single filt. flow)}}{\text{(total run hours)(avg. filt. flow)}}$$

DRF #3 was at maximum flow which was calculated as average run flow × the max/avg. filter flow ratio taken from Figs. 51 and D-1 and D-2, Appendix D.

turbidity, the average effluent turbidity for the alum runs was about 0.08 NTU lower than the average effluent turbidity for the polymer runs. However, the head loss increases for polymer runs were an average of 6 cm/h less than the alum runs as shown in Table 28 (about 75% less than the average CRF head loss increase in the alum runs). Lower head loss increase will give longer filter runs and greater overall water production when the backwash water volume is considered.

Optimum polymer dosages were checked during the preliminary runs by turning off the feeder and observing the subsequent reaction on the continuous recording turbidity graphs as discussed earlier under Filtration Procedures. Runs J-4 and J-9 had changes in optimum polymer dosage during the run as indicated in Table 25.

Comparisons with prior studies--

Constant-rate and declining-rate filtration comparisons have been conducted in pilot and full-scale plants and under varying conditions, with results and conclusions as indicated in the literature review. The most controlled comparison for which significant data are available is the DiBernardo and Cleasby work [26]. This work will be referred to most often in the following discussion.

From general observation of the CRF and DRF effluent turbidity and head loss increase data (Tables 26 and 28) there is no apparent difference in the effluent turbidity and the head loss increase data for the two filtration control methods. Figures 6 and 11 in DiBernardo and Cleasby also show parallel head loss increases on the CRF and DRF [26]. General observation of the particle count and total coliform data indicate that there is no significant difference in the removal of these constituents by CRF and DRF. The data do show the increased turbidity, particle count and total coliform count during the initial improvement period on DRF #3. Typically, DRF #3 filter began operation at 1.6 times average flow for a few minutes just after it was backwashed (i.e., at level 4).

DiBernardo and Cleasby concluded that "the DRF system produced an average filtrate turbidity which was consistently and substantially better than that of the CRF" at three filtration rates. The current study found no difference in the filtrate turbidity. Several hypotheses can be advanced for this conflict of conclusions.

First, the CRF effluent turbidity curve in the prior study, when filtering lime softened water, generally exhibited a gradually deteriorating turbidity after the initial improvement period. No such deterioration was evident in the current study. Thus, the superior filtrate turbidity of the DRF in the prior study may be due to the reduction in rate of filtration as the run progressed, which relaxed the hydraulic shear forces on the filter at the time when deterioration might otherwise have been expected. No such benefit would be expected in the current study because no terminal deterioration was observed, even on the CRF.

Secondly, the prior study imposed some backwash simulations during the operation of the CRF. The spikes in the effluent turbidity caused by these backwash simulations were included in the CRF turbidity average. In contrast, no backwash simulations were used in the CRF of the current study.

Similarly to the previous study, the poorest effluent quality of the declining rate system in the current study occurred after each backwash operation. DiBernardo and Cleasby found that the poorest effluent quality of the CRF occurred during the backwash simulations. However, this conclusion was drawn from a single CRF filter effluent rather than the actual combined effluent from four CRF filters.

Tate et al. reported the one study found in the literature where particle analysis was given for raw and filter effluent in a direct filtration study [82]. Their raw water had 1518 particles/mL in the 2.5 to 150 μm range. They were using a 150 μm sensor whereas a 60 μm sensor was used for the current study. Their raw water count was about 1 log cycle lower than the value shown at 2.5 μm in Figs. 52 and 53 for Run J-6. Tate et al. indicated that everything greater than 10 μm was removed, leaving 8 particles/mL between 2.5 and 10 μm. This was about a 2.25 log reduction. In contrast, for Run J-6 of the current study, the 3 effluents had an average of about 500 particles/mL greater than 2.5 μm which was about a 1.5 log reduction, and about 25 particles/mL greater than 10μm. Thus, the results of the filtration particle count removal efficiency of the current study are poorer than those of the Tate study [82]. A possible explanation for the better filtration performance in the Tate study is that they used flocculation followed by filtration through 20 inches of 1.1 mm es anthracite and 20 inches of 0.5 mm es sand. The current study included no flocculation and utilized a thinner layer of larger coal (14 inches of 1.4 mm es).

Figures 52 and 53 also show the close relationship between CRF and DRF particle count values. The first-hour DRF #3 values are higher than either the DRF or CRF values because it was operating at higher than average rate.

A significant advantage of DRF over CRF is that less total available head loss is required in the filter plant [15,18,26,39,40,72]. Cleasby and DiBernardo [18] gave very detailed calculations to illustrate this advantage by comparing the design of an influent-flow-splitting CRF with a DRF system. The rate of head loss increase was the same for the CRF and the DRF when operated in parallel under identical conditions in DiBernardo and Cleasby and in the current studies. Therefore, the same advantage could be demonstrated from the head loss data of the current study.

DiBernardo and Cleasby found that the highest flow was in the cleanest of the 3 filters not being backwashed [26]. Figures 51 and D-1 through D-3 (Appendix D) show that the highest flow rate was in the cleanest filter just after backwash (an average of 1.60 times the average flow rate). The DiBernardo and Cleasby study and the current studies utilized about a 10-min backwash period. The reason for this difference is not easily understood. It is affected by: (i) the rise in level in the operating filters which occurs during the backwash; (ii) the equilibrium water level that results when the filter inlet of newly backwashed filter is opened, allowing free communication of flow between all of the filters; and (iii) the clean bed water level vs filtration rate curve. It is presumed that the turbulent head loss that existed in the earlier study was higher, reducing the starting rate of filtration through the clean filter.

It should be remembered, however, as Hudson has indicated [40], that the DRF water level rise during backwash would be less than that in the CRF because the dirtiest filter, at 20-40% below average flow, is backwashed as opposed to a filter at average flow in a CRF filter bank. Also, in the CRF system, since there is no free communication between filters, the post backwash flow transitions would occur abruptly compared to the gradual (30-40 min) transitions seen in Figs. 51 and D-1 through D-3.

RESULTS OF PHASE III, DIRECT VS DIRECT IN-LINE FILTRATION

Filter Run Summary

The basic objective of the Phase III investigation was to determine whether there was any difference in the quality of effluents or the head loss development for filters treating flocculated water and unflocculated water.

Pilot plant runs with with naturally occurring lake water were conducted using two direct filtration systems, one with volume flocculation and the other without volume flocculation (i.e., in-line filtration). The removal of turbidity, particles in the range of 7-12 μm and 1-60 μm, fecal coliform bacteria, and the head loss development by both filtration systems was determined.

The pilot plant experiments conducted during Phase III were started on April 19, 1983 and were ended on July 15, 1983. Three different chemical pretreatment schemes were investigated. Efforts were made to carry out runs with each pretreatment at different raw water temperatures and at two filtration rates. A summary of the runs conducted indicating date, filtration rate, chemical dosages, raw water temperature and turbidity, and pH are given in Table 34. More detailed tables on some parameters are presented later. Table 34 excludes the trial filter runs which were conducted with variable coagulant dosage to select the optimum chemical dosage for the subsequent filter run. The runs in Table 34 include 16 formal runs (run numbers without letters appended) in which all parameters were measured. The other runs presented in Table 34 were either trial runs or replicating runs in which only head loss and turbidity were measured. The experiments began in the spring with a water temperature of 7° C and ended in midsummer with a water temperature of 22° C. It was hoped to include periods of serious algal bloom in Phase III, as had been experienced the previous summer. However, a cool early summer delayed the development of warm water and the experiments were terminated, due to budget constraints, before any major algal blooms had occurred.

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Several operational changes made during this phase are listed chronologically below:

April 19, 1983	Started operation of Phase III.
April 23	Reduced turbine blades in each flocculator cell from 6 to 3 to reduce motor load.
April 23	Calibrated Ratio Turbidimeter with formazin suspension.
April 29	Improved control of filtration rate by installing second splitter box on flocculator effluent.
May 10	Calibrated Ratio Turbidimeter with formazin suspension.
May 12	Revised effluent piping of both filters to reduce initial head loss.
May 13	Installed larger inlet pipe to filter 2 to correct overflow problems at high filtration rates.
May 16	Switched from reagent grade alum to commercial grade alum.
May 24	Provided mixing in sewage tank with submersible pump.

TABLE 34. SUMMARY OF PILOT PLANT EXPERIMENTS IN PHASE III USING ALUM OR CAT-FLOC T

Run	Date 1983	Rate (m/h)	Chemicals (mg/L)		Acid Used	Raw Water		Water pH	
			Alum	Polymer		Temp. (°C)	Avg. Turb. (NTU)	Raw	Filtered
L-1	4/20-22	7.0	5.5		--	7	3.5	8.3	8.2
L-2a	4/22-25	6.8	8.3		yes	9	3.7	8.3	7.1
L-2	4/25-27	6.8	8.3		yes	9	3.1	8.3	7.0
L-3a	4/29-5/2	6.8		Various*		10	3.3	8.4	8.4
L-3	5/2-6	7.8		0.25*	--	10.5	3.3	8.4	8.3
M-1	5/9-12	7.8	4.9		yes	12	3.2	8.4	6.8
M-1b	5/13	16.1	6.0		yes	14	3.0	8.3	6.8
M-1c	5/14	15.9	6.6		--	15	2.9	8.4	8.3
M-2	5/16-17	16.1	6.1		--	15	2.5	8.5	8.2
M-2a	5/17-18	15.9	6.2		--	13.5	3.3	8.3	8.2
M-3	5/18-19	15.9	6.0		yes	13	2.7	8.3	6.7
M-3a	5/19-20	15.9	5.9		yes	13	3.9	8.3	6.7
M-3b	5/21-22	16.1	7.2		yes	14	2.6	8.3	6.7
M-4a	5/23-24	14.9		Various*		15	2.3	8.3	8.3

TABLE 34. CONTINUED

Run	Date 1983	Rate (m/h)	Chemicals (mg/L)		Acid Used	Raw Water		Water pH	
			Alum	Polymer		Temp. (°C)	Avg. Turb. (NTU)	Raw	Filtered
M-4	5/24-25	14.9		0.45*	--	16	2.3	8.4	8.4
M-4b	5/26-28	15.1		0.44*	--	17	4.8	8.4	8.3
M-5	5/30-31	15.1		3.1†	--	16	8.0	8.4	8.4
N-1a	6/1-5	7.6		0.42*	--	17	11.0	8.4	8.4
N-2a	6/6	7.6	Various			18.5	9.6	8.3	8.2
N-2	6/6-8	7.6	6.7		--	19	7.1	8.2	8.0
N-2b	6/9-11	7.6	5.7		--	20	6.8	8.1	8.0
N-3a	6/11-13	12.7	5.9		--	21.5	4.9	8.2	8.1
N-3	6/13-14	14.4	6.4		--	22	6.8	8.2	7.9
N-4	6/15-16	14.4	5.9		--	22.5	8.3	8.1	7.9
N-4a	6/16-18	14.4	4.0		--	22	2.1	8.1	8.0
N-4b	6/18-20	14.4	5.1		--	21.5	1.9	8.1	7.9
O-1b	6/22-24	14.7		0.35*	--	19	3.7	8.1	8.1
O-1c	6/25-27	14.9		0.38*	--	19.5	3.2	8.0	8.0

TABLE 34. CONTINUED

Run	Date 1983	Rate (m/h)	Chemicals (mg/L)		Acid Used	Raw Water		Water pH	
			Alum	Polymer		Temp. (°C)	Avg. Turb. (NTU)	Raw	Filtered
O-1	6/27-28	15.1		0.38*	--	19	9.1	7.6	7.6
O-2	6/28-29	15.1		0.38*	--	19.5	10.5	7.5	7.5
O-3a	6/30-7/4	7.8		0.35*	--	21	7.5	7.7	7.7
O-3	7/5-7	7.8		0.35*	--	21	7.5	7.9	7.9
P-1a	7/9-11	7.8				21	3.5	7.9	7.0
P-1	7/11-13	7.8	4.6		yes	21.5	2.0	7.8	6.7
P-2	7/13-14	15.1	4.5		yes	21.5	2.9	7.7	6.7
P-2a	7/14-15	13.9	4.5		yes	21.5	2.4	7.6	6.7

[*] Cat-Floc T

[†] Culligan F-86

May 25	Installed bubble trap on raw water turbidimeter to correct developing air bubble problem due to leaking pump seal.
June 1	Began use of Cat-Floc T dated 7/82. Prior to this date Cat-Floc T dated 6/81 was used.
June 7	Replaced submersible pump in sewage tank with 2-bladed stirrer to reduce heating of the sewage.
June 15	Received and used new Cat-Floc T supply in all subsequent polymer runs.
June 22	Moved raw water intake to about 30 m from lake shore at 6 m depth. Prior location was near shore at about 2 m depth resulting in erratic raw water turbidity on windy days due to bank erosion.
June 28	Noted failure of no. 3 cell flocculator drive; out of service until July 11.
July 12	Noted algae bloom occurring.
July 15	Ended operation of Phase III.

TYPICAL FILTER RUN DATA, PHASE III

Runs L-1, M-3, M-4 and O-3 will be used to illustrate the data collected through each formal run. For run L-1, alum was the primary coagulant used at a raw water temperature of 7° C; for M-3, alum and acid were used, at 13° C; and for M-4 and O-3, Cat-Floc T was used at 16° C and 21° C, respectively. Turbidity results for these four runs are shown in Figs. 54 through 57. The results given in these figures for influent turbidity correspond to the raw water turbidity prior to the addition of chemicals and sewage.

Breakthrough was a common characteristic observed in the runs that were carried out using alum as primary coagulant, regardless of the pH. In contrast, in those runs using Cat-Floc T, terminal head loss was reached before any breakthrough trend was detected.

Figures 58 and 59 show the first hour of filtration for runs N-4 and O-1, using alum and Cat-Floc T, respectively. The filtration rate was similar for both runs. A sharp reduction in turbidity was consistently observed within that period for runs using alum; a more gradual reduction was observed when Cat-Floc T was used. Filter 2, which received flocculated water, produced a better quality filtrate in the first hour than Filter 3, especially when alum was used; for Cat-Floc T in several instances, the opposite was true.

Particle count results in the 7-12 μm size range for runs L-1, M-3, M-4 and O-3 are shown in Figs. 60 through 63. As the results shown for particle

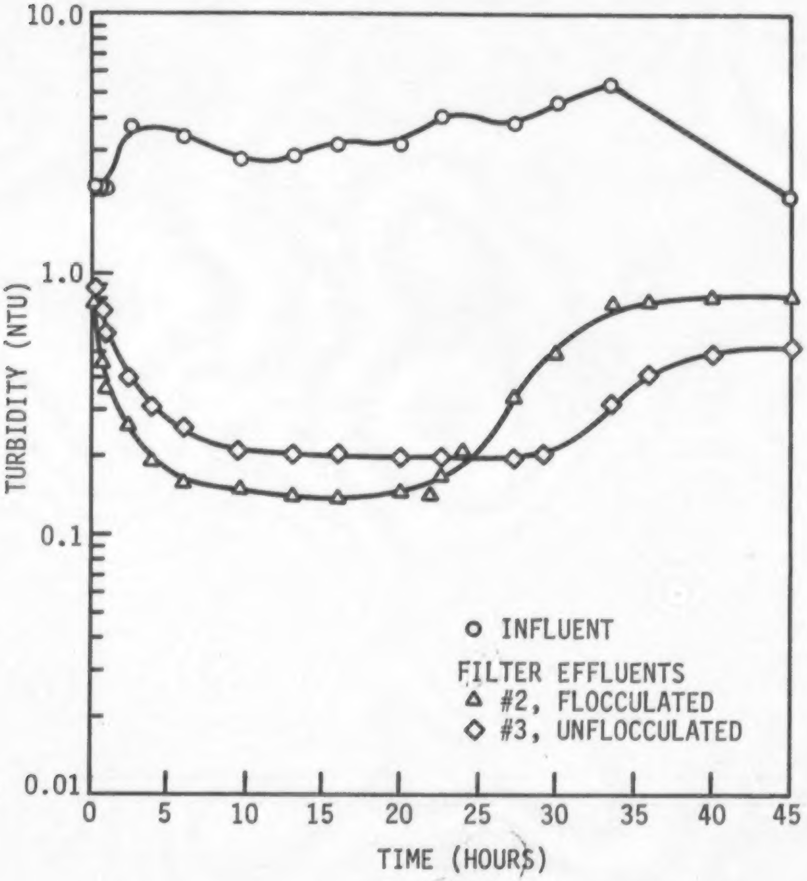

Fig. 54. Turbidity for run L-1 at 7.0 m/h (2.86 gpm/ft^2) using alum.

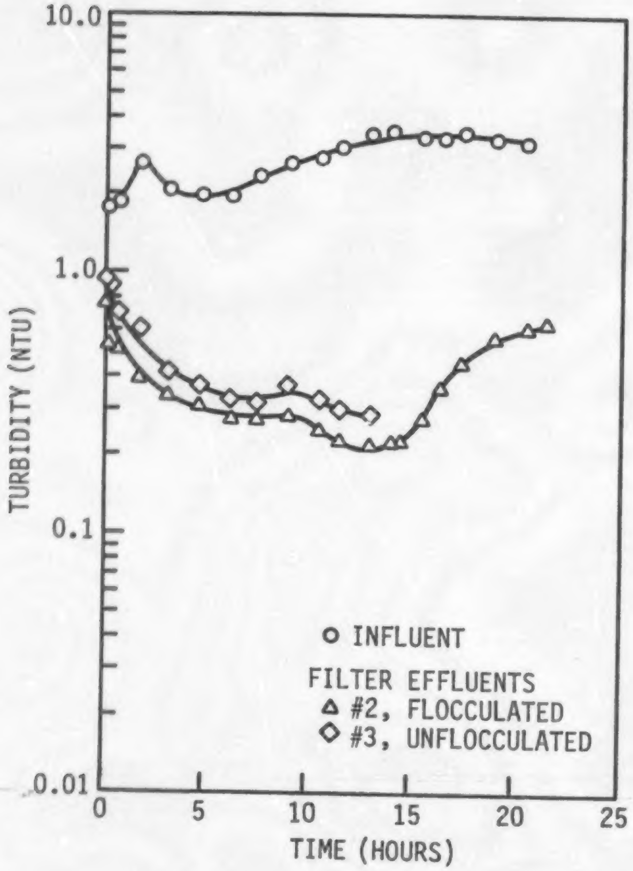

Fig. 55. Turbidity for run M-3 at 15.9 m/h (6.5 gpm/ft^2) using alum and acid.

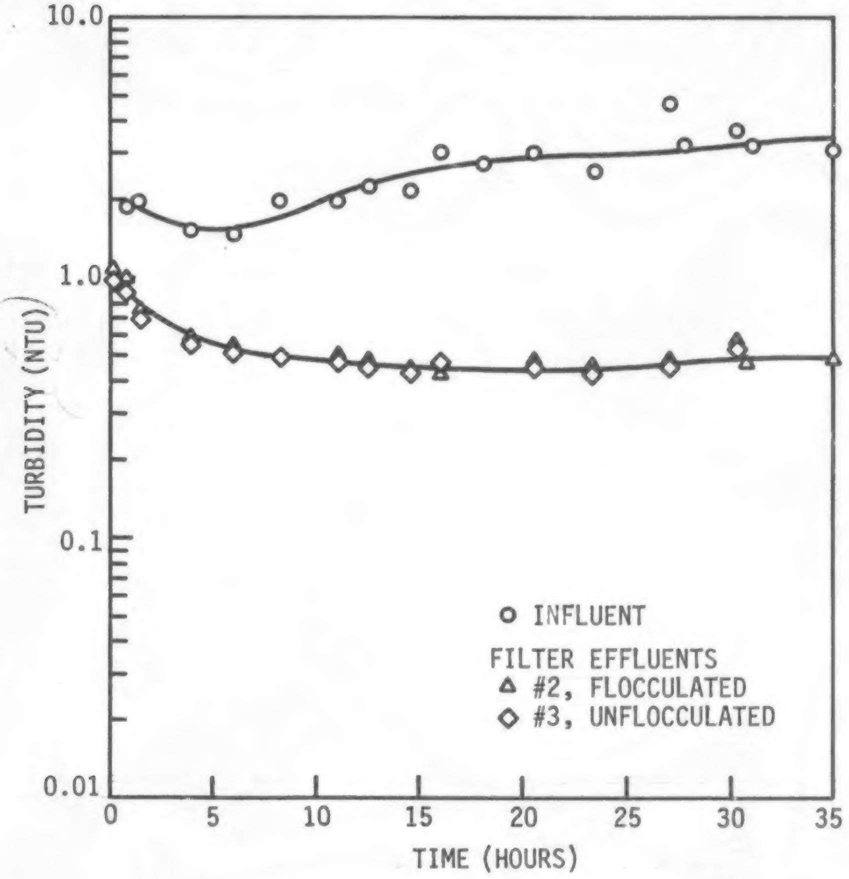

Fig. 56. Turbidity for run M-4 at 14.9 m/h (6.1 gpm/ft^2) using Cat-Floc T.

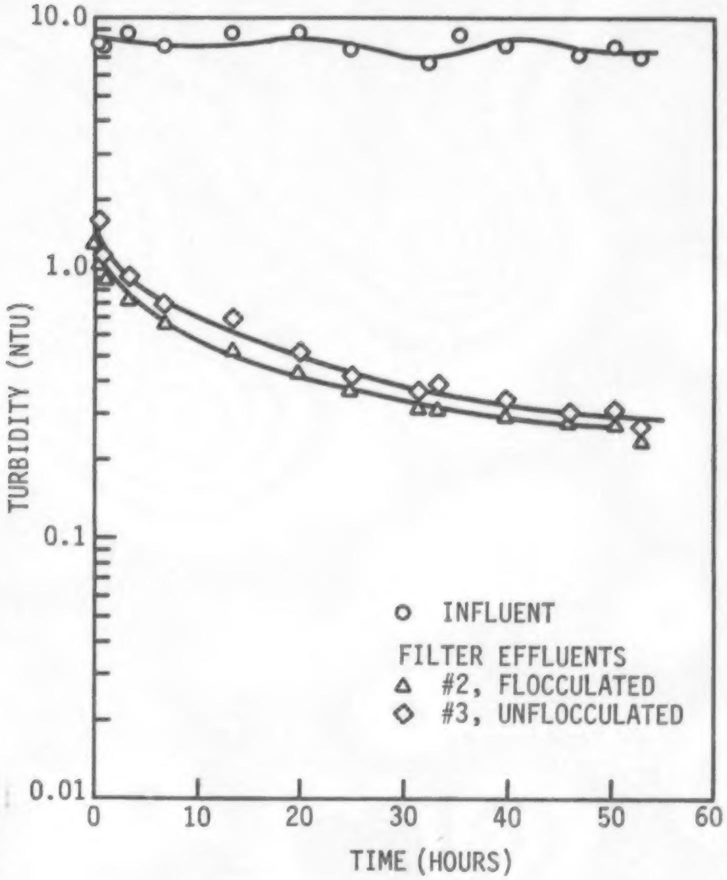

Fig. 57. Turbidity for run O-3 at 7.8 m/h (3.2 gpm/ft^2) using Cat-Floc T.

Cat-Floc T

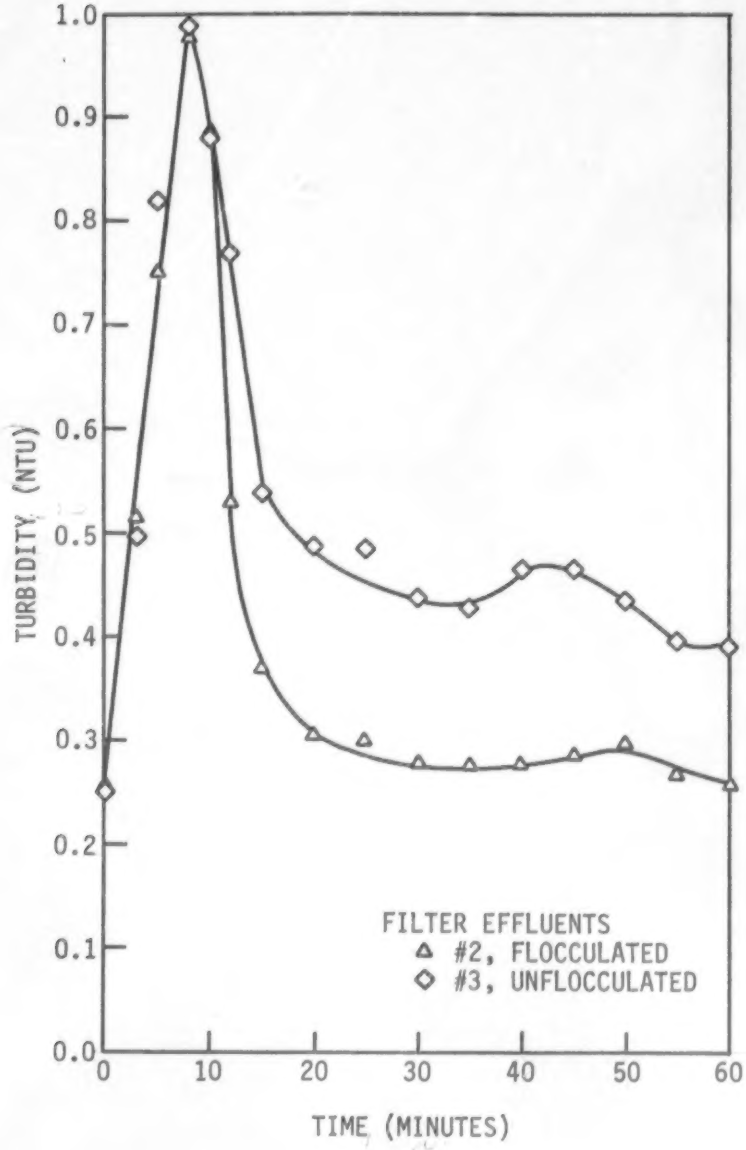

Fig. 58. Effluent turbidities during the first hour of run for run N-4 at 14.4 m/h (5.9 gpm/ft^2) using alum.

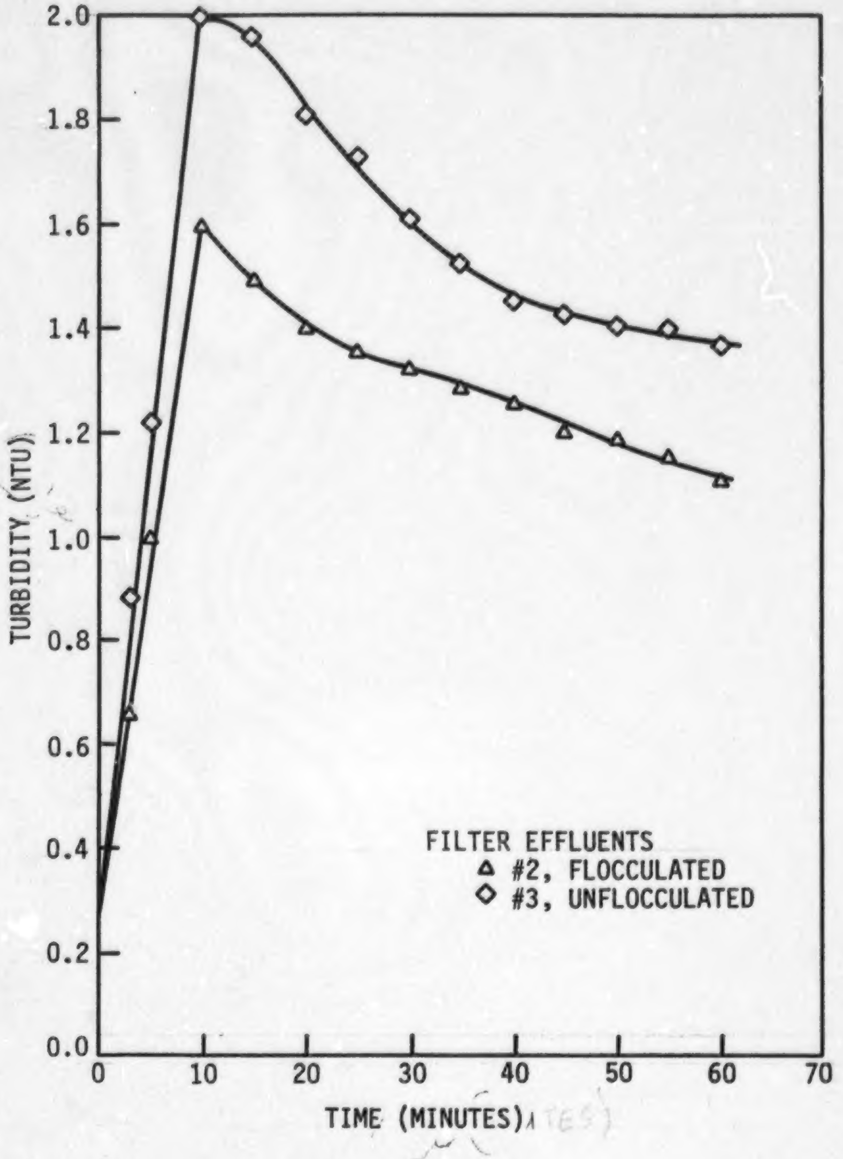

Fig. 59. Effluent turbidities during the first hour of run for run O-3 at 7.8 m/h (3.2 gpm/ft^2) using Cat-Floc T.

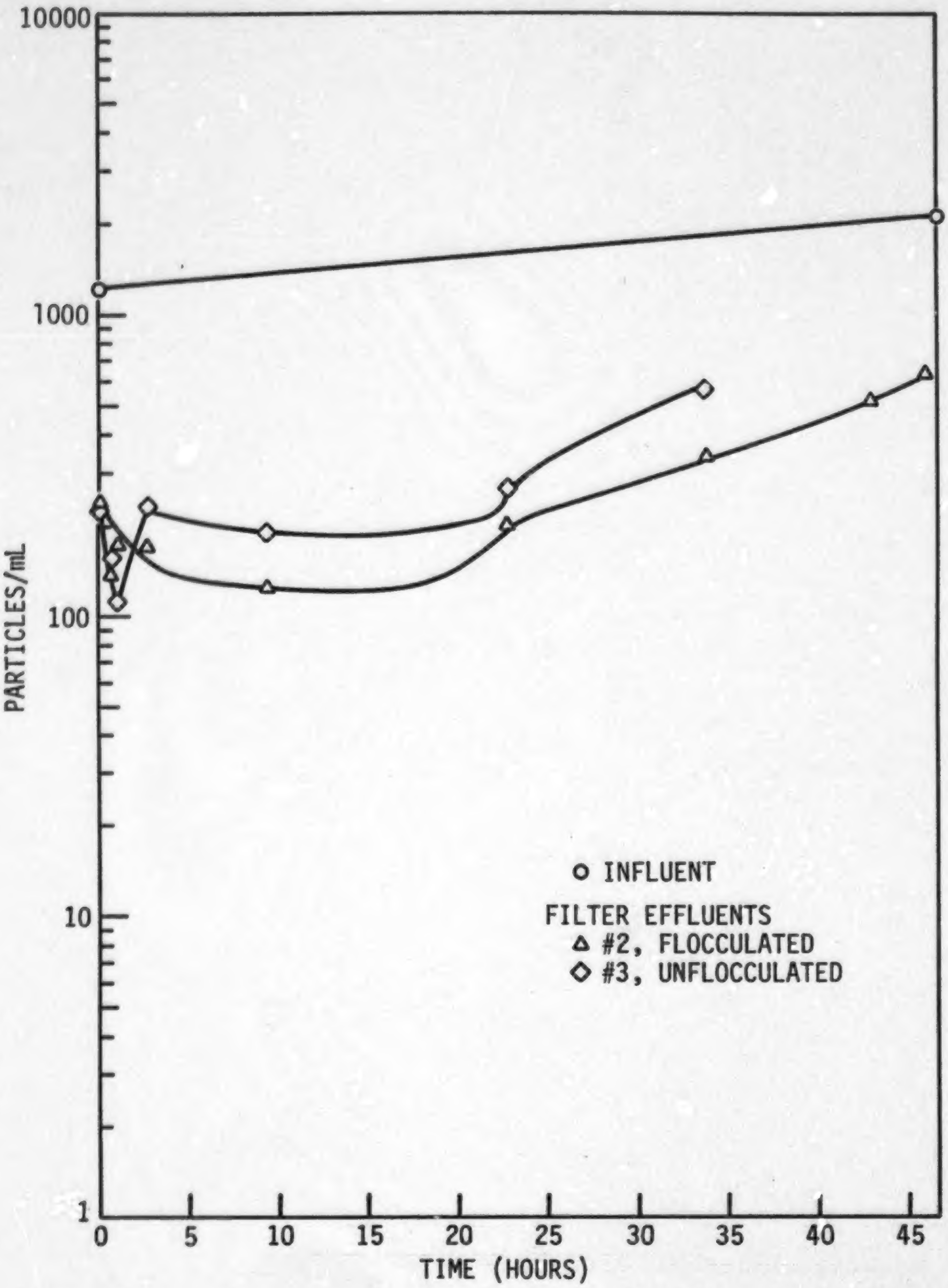

Fig. 60. 7-12 μm particle count for run L-1 at 7.0 m/h (2.85 gpm/ft^2) using alum.

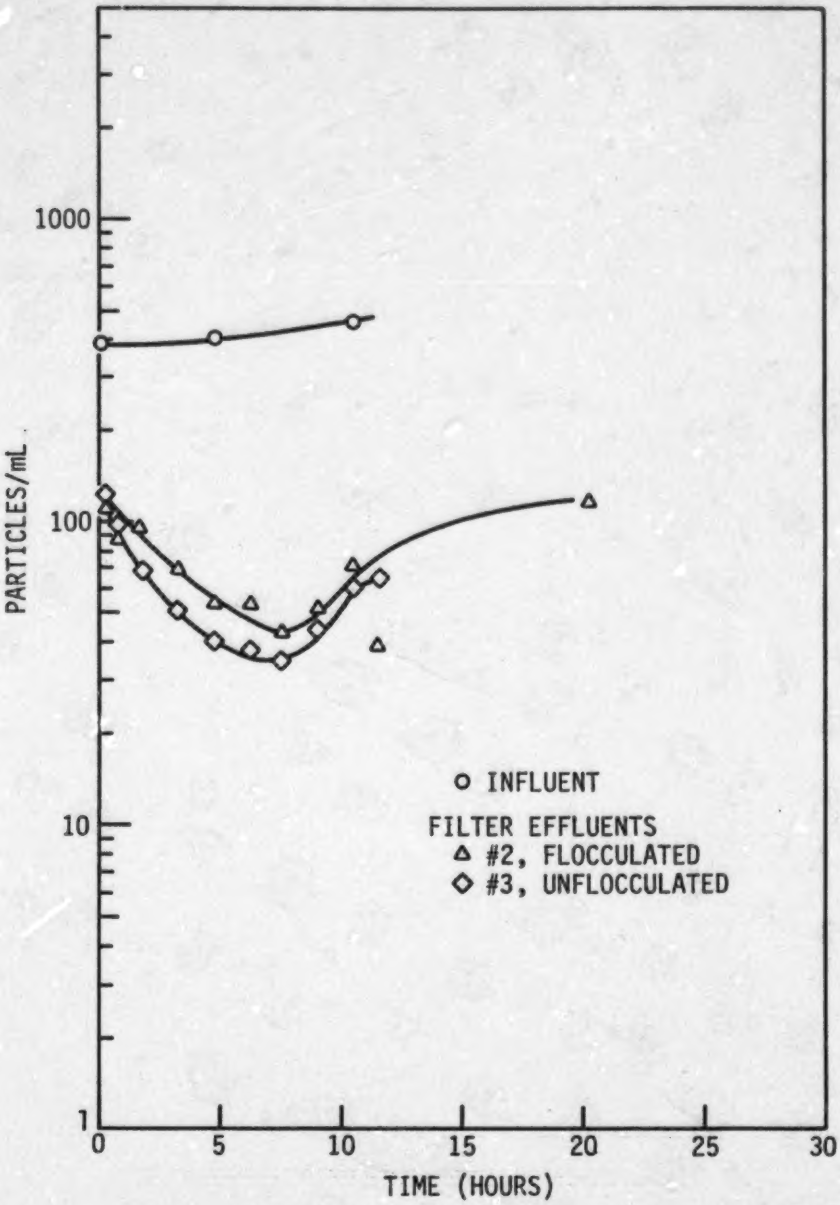

Fig. 61. 7-12 μm particle count for run M-3 at 15.9 m/h (6.5 gpm/ft^2) using alum and acid.

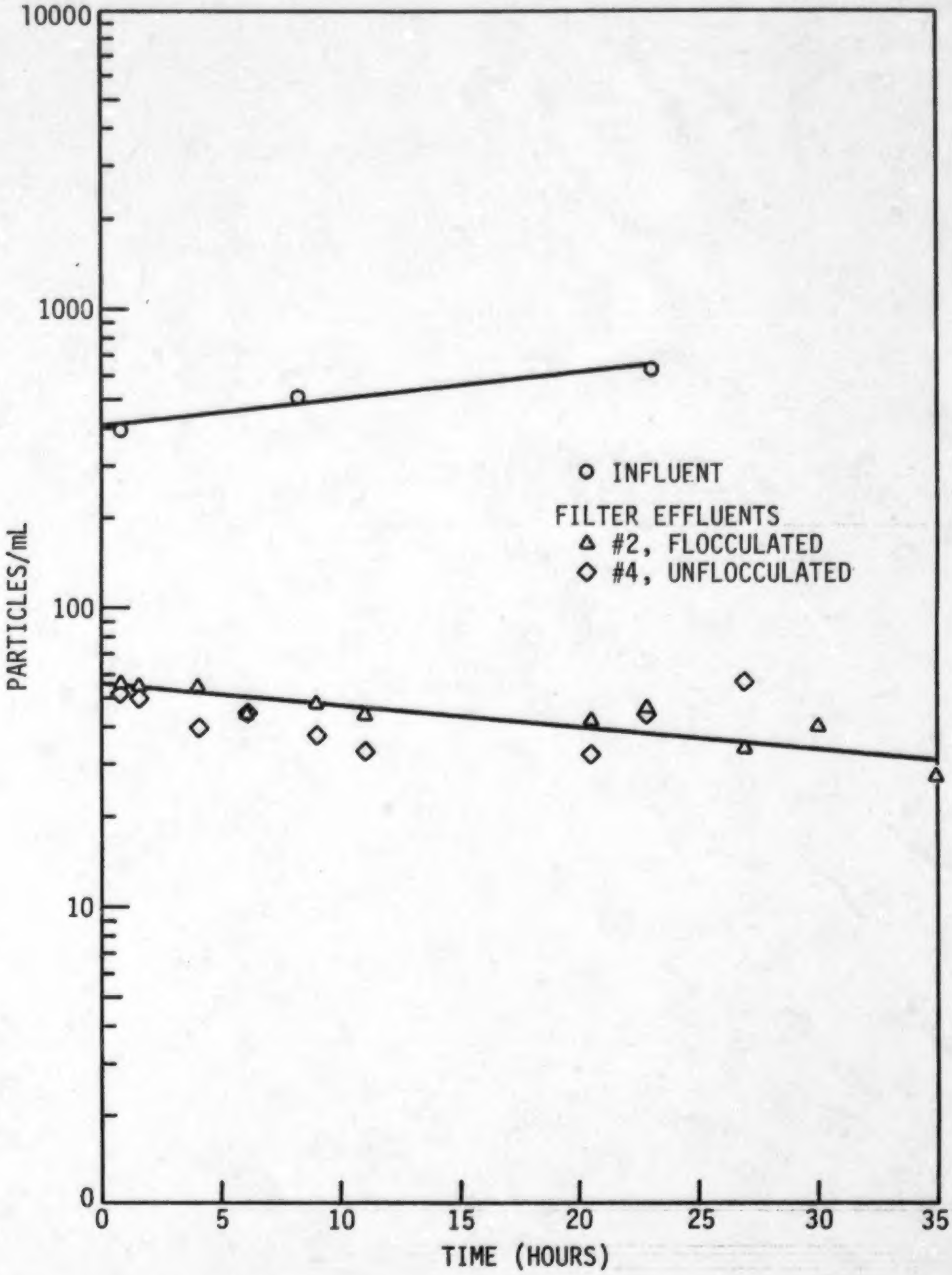

Fig. 62. 7-12 μm particle count for run M-4 at 14.9 m/h (6.1 gpm/ft^2) using Cat-Floc T.

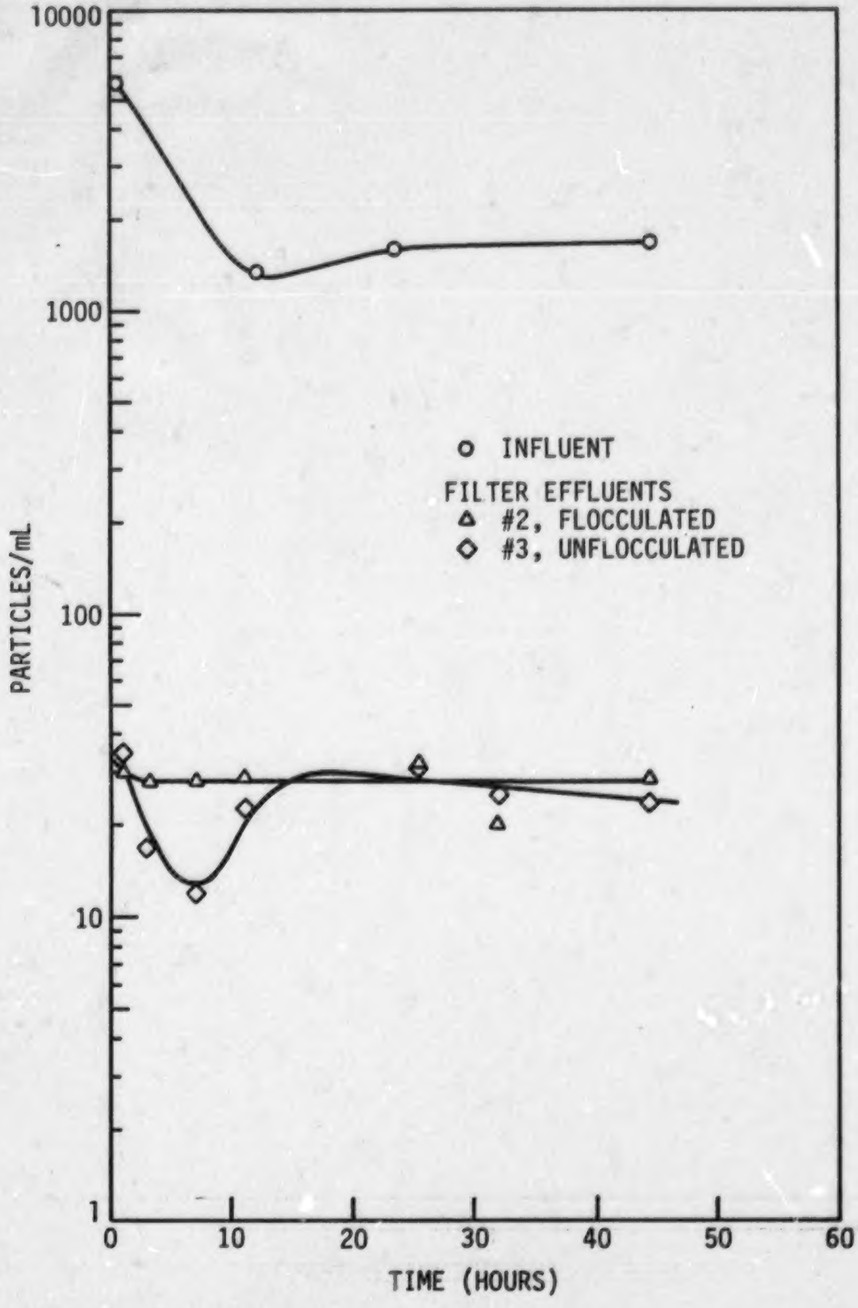

Fig. 63. 7-12 μm particle count for run O-3 at 7.8 m/h (3.2 gpm/ft^2) using Cat-Floc T.

count are data points based on grab samples at various times, the profile of this variable against time is not as smooth as the profile for turbidity, which was continuously recorded. The lines through the data points were visually drawn, the shape being guided by the turbidity graphs for the same filter run. The turbidity breakthrough behavior in the runs using alum or alum and acid is supported by similar breakthrough evident in Figs. 60 and 61. Runs M-4 and O-3, in which Cat-Floc T was used, show a consistent particle removal behavior with no signs of breakthrough, regardless of the flow rate used in this study.

Figures 64 through 67 show the head loss accumulation data for Runs L-1, M-3, M-4 and O-3. The final water level is not the same in every graph since overflow was not always reached. The difference in initial head loss evident between L-1 and O-3, both at the same nominal filtration rate, was because of the change in effluent piping on May 12. A significant difference was observed in the behavior of head loss accumulation between the runs carried out with alum and those carried out with Cat-Floc T. Figures 64 and 65 show nearly straight-line-type head loss accumulation, characteristic of deep-bed solids removal, whereas Figs. 17 and 18 show some exponential-type head loss accumulation, characteristic of partial removal of solids in the top layer of the filter.

The bacterial data collected during Phase III are not presented because of their erratic behavior. Some of the problems are discussed below.

a) In the early runs of Phase III, samples were analyzed for total coliform count. However, an erratic and unpredictable level of total coliforms in the influent samples sometimes resulted in the use of inappropriate dilution factors. Also, the membranes were often covered with competing growth colonies that suppressed the growth of coliform colonies. Oftentimes, coliform colonies were only evident around the edge of the growth on the membrane.

b) To correct these problems, and after some trials running both total coliform and fecal coliform counts on the same samples, it was decided to switch to fecal coliform analysis.

c) Some improvement occurred, but the influent counts seemed to decrease with time during a filter run due to sedimentation in the sewage tank. Therefore, on May 24, 1983, a submersible pump was installed within the sewage tank to maintain a homogeneous bacterial suspension. However, the pump motor heated the remaining sewage up to 39° C (102° F) during the run, which seemed to promote bacterial growth up to four to six times the initial number. Thus, the number of coliform bacteria being fed to the filters increased continuously during the filter run.

d) After installing a paddle mixer in place of the submersible pump, the number of coliforms being fed stabilized. About that time, however, heavy and persistent rainfall and sewer infiltration diluted the raw sewage coming to the Ames Wastewater Treatment Plant and the fecal coliform tests were plagued by low fecal coliform counts and

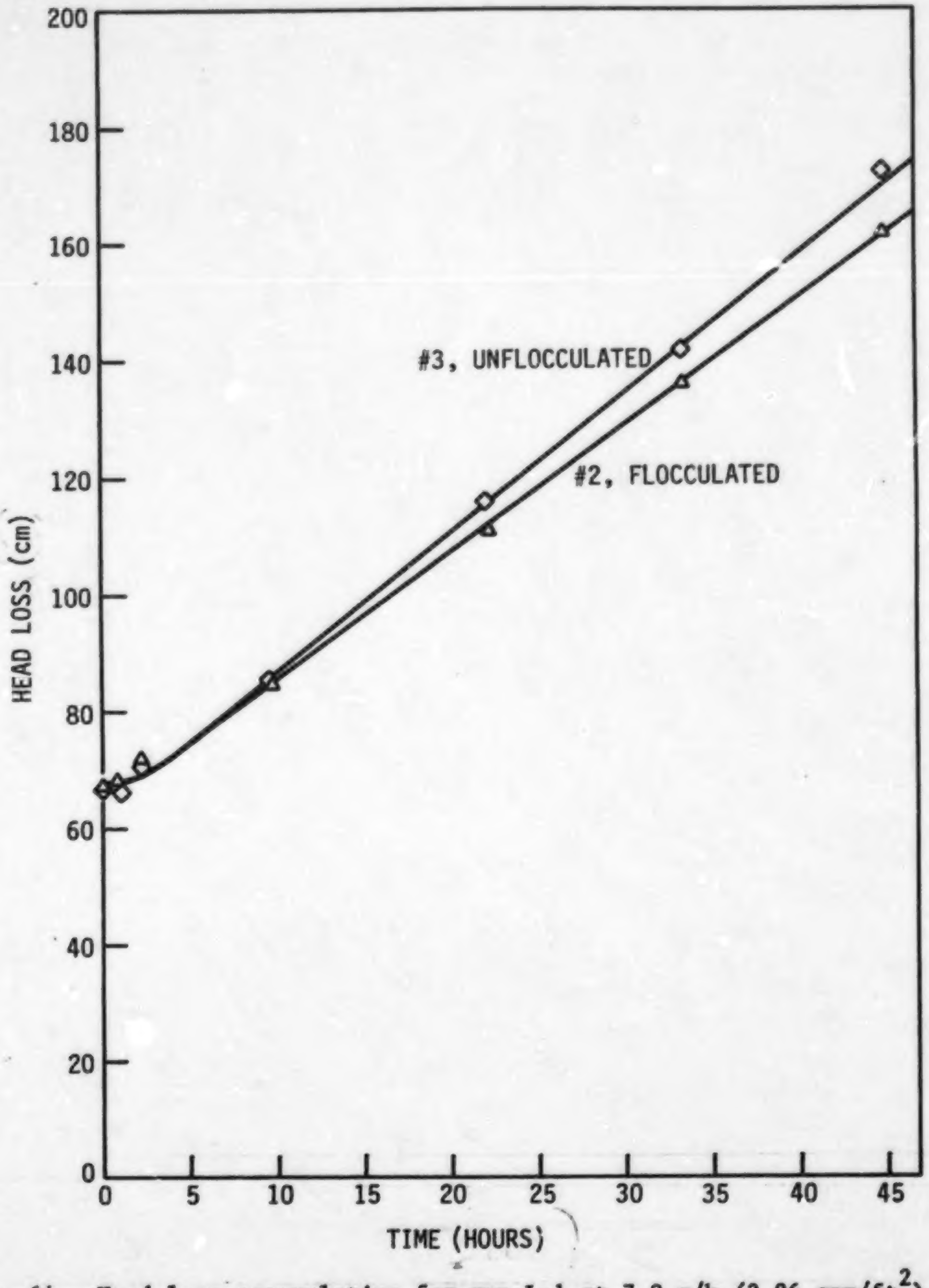

Fig. 64. Head loss accumulation for run L-1 at 7.0 m/h (2.86 gpm/ft^2) using alum.

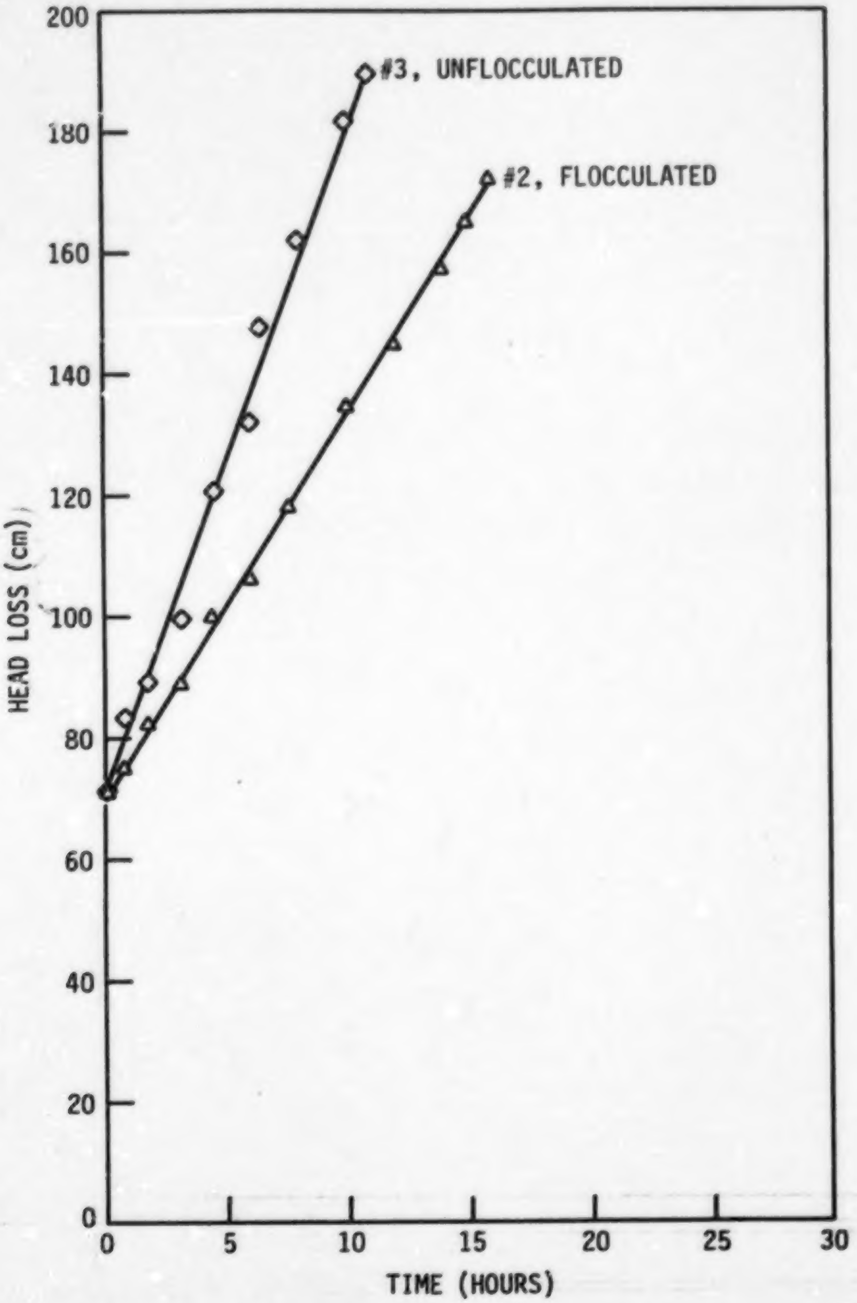

Fig. 65. Head loss accumulation for run M-3 at 15.9 m/h (6.5 gpm/ft^2) using alum and acid.

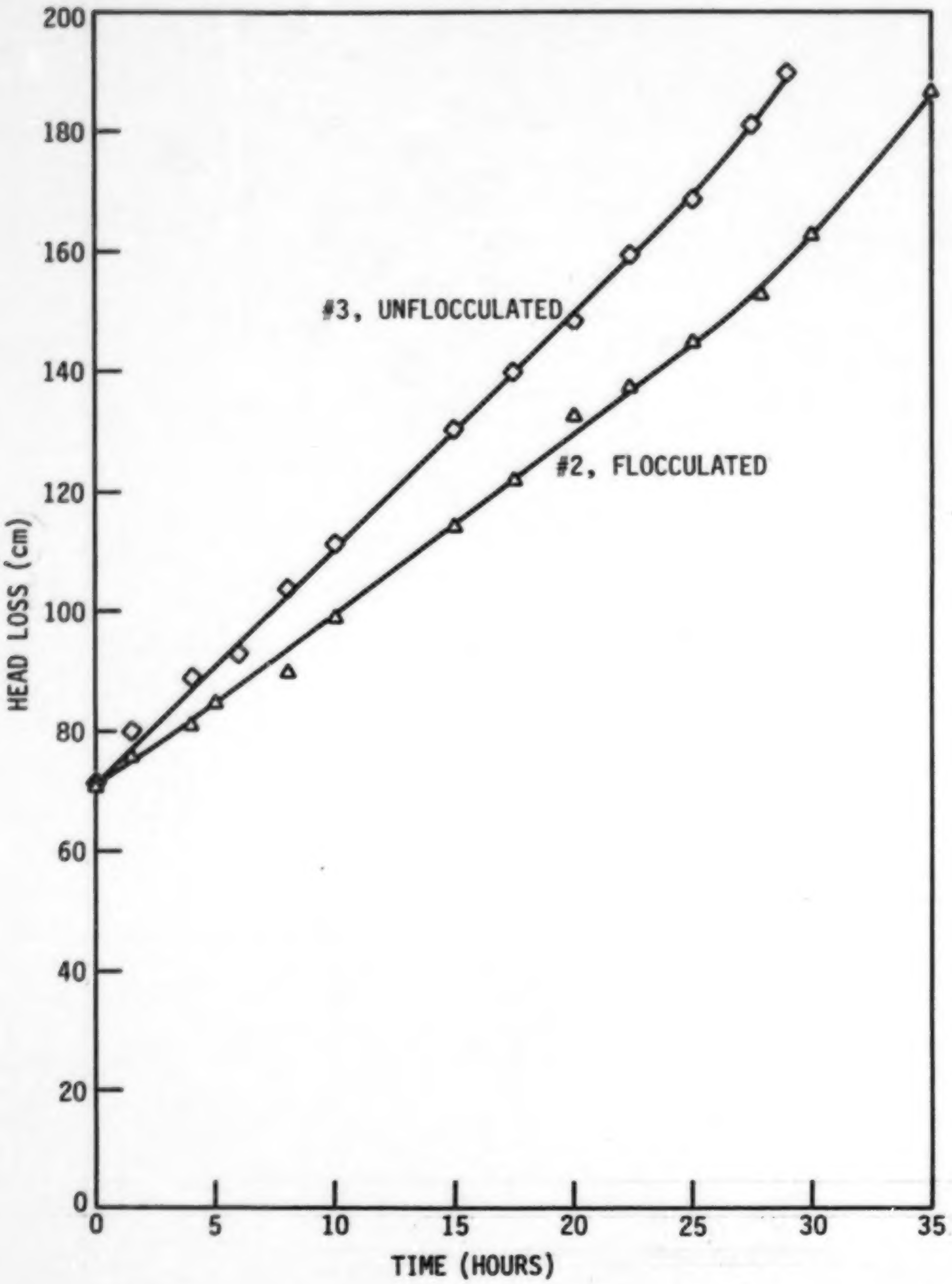

Fig. 66. Head loss accumulation for run M-4 at 14.9 m/h (6.1 gpm/ft^2) using Cat-Floc T.

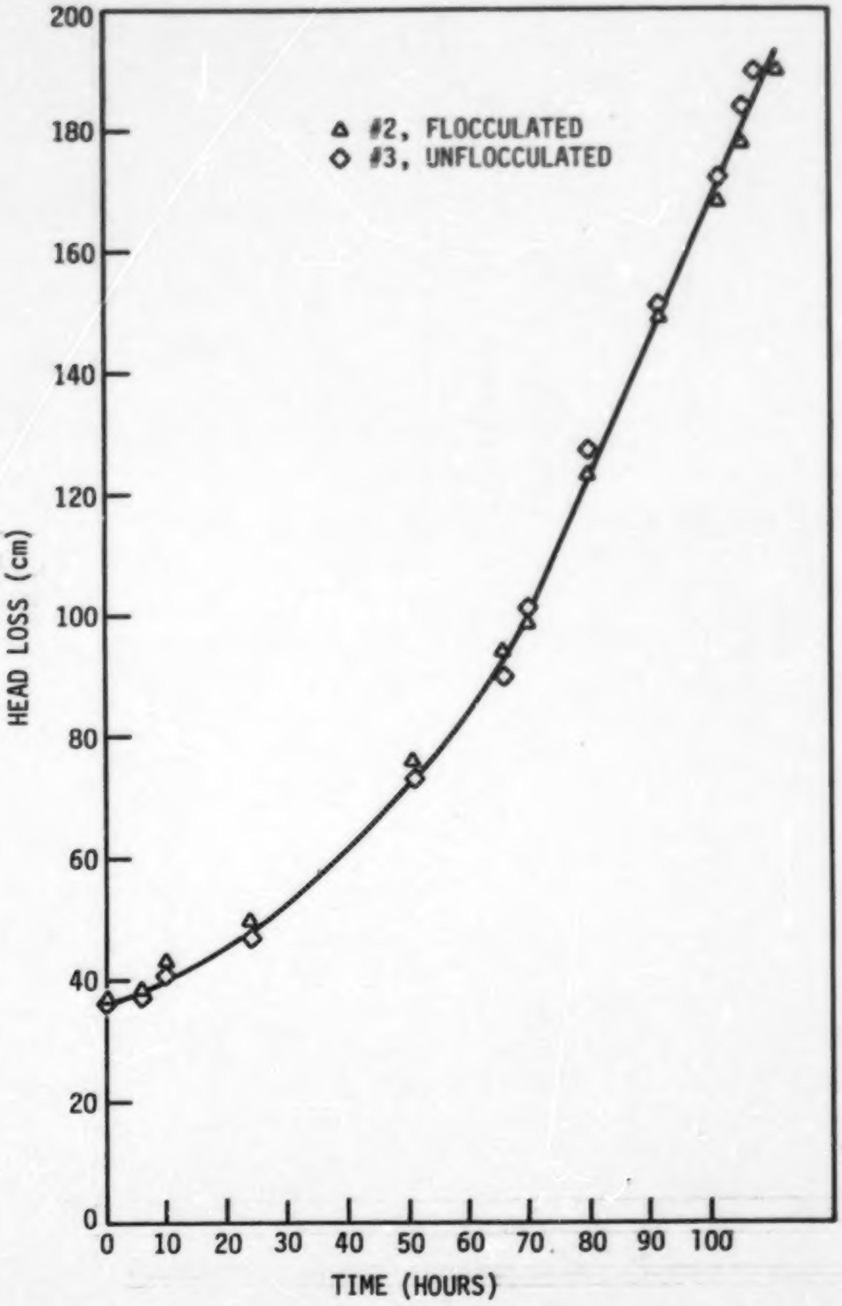

Fig. 67. Head loss accumulation for run O-3 at 7.8 m/h (3.2 gpm/ft^2) using Cat-Floc T.

by competing growth of atypical colonies, which prevented the proper enumeration of fecal coliform bacteria.

TABULAR RESULTS SUMMARY, PHASE III

Pictorial results of Runs L-1, M-3, M-4 and O-3 were presented in previous pages. It was intended in those figures to show the performance trends of the two filters that were consistently observed. However, not 4 but 16 formal runs were monitored for turbidity, particle count and head loss accumulation. In addition, turbidity and head loss data were available for 20 trial and duplicate runs. The data from the 16 formal runs as well as from the trial and duplicate runs are summarized in a series of tables from which comparison of several parameters will be made. These comparisons will be carried out taking into account the primary coagulant that was used, the filtration rate, and the pH. The parameters compared are turbidity, particle count, and head loss accumulation.

Turbidity and Filter Run Length

As a continuous record of influent and effluent turbidity was obtained, an average turbidity was determined for each of the three identifiable periods in the runs carried out using alum: initial improvement period, rest of the run prior to breakthrough, and breakthrough period. For the runs carried out using Cat-Floc T, and for some alum runs, breakthrough was not observed so the average values were determined only for the first two periods mentioned.

The average turbidity for every period was determined by graphical integration of the area under the turbidity vs time curve. For the first hour of run, the area under the curve of turbidity vs time was divided into four spaces and thus integrated. For the remainder of the run, different time divisions were employed as the length of the runs varied substantially from one to another. When integrating the curve for the breakthrough period, different time divisions were also made, depending on both the length of the breakthrough period observed and the slope of the turbidity curve during that period.

The turbidity results for 36 runs are shown in Table 35. The maximum, average, and low turbidity values for raw water during the run are also given. As seen in Table 35, there were significant differences in raw water quality and also between the results obtained for runs carried out using alum and runs using cationic polymer. However, attention should be focused on the differences between Filter 2 processing flocculated water and Filter 3 processing unflocculated water. To assist in that comparison, the Filter 3/Filter 2 ratio of the individual values is presented in Table 36.

From the raw turbidity averages of Table 35 and the ratio values of Table 36, the following observations can be made.

Lower turbidities were consistently achieved prior to breakthrough with Filter 2 than with Filter 3 when alum was used, regardless of the pH. For

TABLE 35. SUMMARY OF RAW AND AVERAGE FILTER EFFLUENT TURBIDITIES DURING ALL FILTER RUNS

Run Number	Water Temp. ° C	Raw Turbidity (NTU)			Average Filter Effluent Turbidity (NTU)					
					F2, with Flocculation			F3, Unflocculated		
		High	Avg.	Low	1st h	Remainder	Breakthrough	1st h	Remainder	Breakthrough
Alum Runs at 7.5 m/h (nominally)										
L-1	7.0	5	3.5	2	0.47	0.17	0.71(12)*	0.84	0.24	0.41(7)*
N-2a	18.5	30	9.6	1.5	0.51	--	--	0.51	--	--
N-2	19.0	20	7.1	3	0.36	0.22	0.84(13)	0.71	0.34	0.68(11)
N-2b	20	12	6.8	4	0.33	0.21	0.42(8)	0.42	0.30	0.38(7)
				$\bar{x}$ =	0.42	0.20		0.62	0.29	
Alum Runs at 15 m/h (nominally)										
M-1c	15	4	2.9	2	0.31	0.17	0.78(6)	0.59	0.26	No
M-2	16.1	5	2.5	1	0.54	0.23	0.34(5)	0.55	0.29	No
M-2a	13.5	4	3.3	3	0.55	0.30	No	0.68	0.37	No
N-3a	21.5	10	4.9	2	0.41	0.21	0.64(11)	0.58	0.26	0.24(1)
N-3	22	18	6.8	3	0.27	0.24	0.39(2)	0.42	0.32	No
N-4	22.5	14	8.3	6	0.33	0.25	0.84(12)	0.63	0.33	0.42(8)
N-4a	22	3.5	2.1	1.5	0.53	0.23	0.46(21)	0.65	0.29	0.39(21)
N-4b	21.5	2.5	1.9	1.5	0.30	0.21	0.34(10)	0.43	0.29	No
				$\bar{x}$ =	0.40	0.23	--	0.56	0.30	--
Alum + Acid Runs at 715 m/h (nominally)										
L-2a	9	4	3.7	3	0.57	0.20	No	0.89	0.29	No
L-2	9	5	3.1	2	0.71	0.26	0.76(12)	1.20	0.26	0.36(10)
M-1	12	7	2.8	1.5	0.63	0.24	No	0.79	0.28	No

TABLE 35. CONTINUED

Run Number	Water Temp. ° C	Raw Turbidity (NTU)			Average Filter Effluent Turbidity (NTU)					
					F2, with Flocculation			F3, Unflocculated		
		High	Avg.	Low	1st h	Remainder	Breakthrough	1st h	Remainder	Breakthrough
P-1a	21	6	3.5	2	0.76	0.31	0.52(20)	0.92	0.38	0.43(10)
P-1	21.5	3.5	2.0	1.5	0.56	0.36	0.52(11)	0.69	0.49	0.54(9)
				$\bar{x}$ =	0.65	0.27	--	0.96	0.35	--
Alum + Acid at 15 m/h (nominally)										
M-1b	14	5	3.0	1.0	0.51	0.21	0.79(5)	0.74	0.32	No
M-3	13	3	2.7	2	0.62	0.31	0.62(8)	0.97	0.41	No
M-3a	13	5	3.9	3	1.34	0.61	0.78(10)	1.66	0.66	No
M-3b	14	3	2.6	2	0.63	0.29	0.44(6)	1.03	0.33	No
P-2	21.5	4.6	2.9	1.9	1.08	0.59	0.86(10)	1.68	0.66	0.75(10)
P-2a	21.5	4.0	2.4	1.7	1.19	0.51	No	1.60	0.62	No
				$\bar{x}$ =	0.90	0.42	--	1.28	0.50	--
Cat-Floc T Runs at 7.5 m/h (nominally)										
L-3a	10	5	3.3	2	0.77	0.60	No	0.69	0.60	No
L-3	10.5	7	3.3	2	1.38	0.54	No	1.20	0.53	No
N-1a	17	20	11	3	1.04	0.43	No	0.94	0.40	No
O-3	21	12	7.5	6	1.29	0.43	No	1.97	0.60	No
				$\bar{x}$ =	1.12	0.50	--	1.20	0.53	--

TABLE 35. CONTINUED

Run Number	Water Temp. ° C	Raw Turbidity (NTU)			Average Filter Effluent Turbidity (NTU)					
					F2, with Flocculation			F3, Unflocculated		
		High	Avg.	Low	1st h	Remain-der	Break-through	1st h	Remain-der	Break-through
Cat-Floc T Run at 15 m/h (nominally)										
M-4a	15	3	2.3	2	0.93	0.58	No	1.13	0.61	No
M-4	16	3	2.3	2	1.00	0.52	No	0.86	0.51	No
M-4b	17	6	4.8	4	0.91	0.54	No	0.74	0.49	No
O-1b	19	5	3.7	1.7	0.58	0.36	No	0.75	0.45	No
O-1c	19.5	5.0	3.2	1.8	0.70	0.37	No	0.74	0.44	No
O-1	19	15	9.1	4	0.57	0.31	No	0.86	0.39	No
O-2	19.5	15	10.5	8	0.69	0.35	No	0.97	0.46	No
				$\bar{x}$ =	0.76	0.43	--	0.86	0.48	--

*hours of breakthrough observation.

TABLE 36. SUMMARY OF RATIO OF AVERAGE EFFLUENT TURBIDITIES FOR FILTERS RECEIVING UNFLOCCULATED WATER (F3)/FLOCCULATED WATER (F2), DURING THE THREE PERIODS OF FILTRATION RUNS

	Ratio of Average Turbidities F3/F2		
Run Number	First Hour	Remainder	Breakthrough
Alum Runs at 7.5 m/h (nominally)			
L-1	1.79	1.71	0.57
N-2a	1.00	--	--
N-2	1.97	1.54	0.81
N-2b	1.27	1.24	0.90
Alum Runs at 15 m/h (nominally)			
M-1c	1.90	1.53	--
M-2	1.02	1.26	--
M-2a	1.24	1.23	--
N-3a	1.41	1.24	0.38
N-3	1.56	1.33	--
N-4	1.91	1.32	0.50
N-4a	1.23	1.26	0.85
N-4b	1.43	1.38	--
Alum + Acid Runs at 7.5 m/h (nominally)			
L-2a	1.56	1.45	--
L-2	1.69	1.00	0.47
M-1	1.25	1.17	--
P-1a	1.21	1.22	0.83
P-1	1.23	1.36	1.03

TABLE 36. CONTINUED

Run Number	Ratio of Average Turbidities F3/F2 First Hour	Remainder	Breakthrough
Alum + Acid Runs at 15 m/h (nominally)			
M-1b	1.45	1.52	--
M-3	1.56	1.32	--
M-3a	1.24	1.08	--
M-3b	1.63	1.14	--
P-2	1.56	1.12	0.87
P-2a	1.34	1.22	--
Cat-Floc T Runs at 7.5 m/h (nominally)			
L-3a	0.90	1.00	--
L-3	0.87	0.98	--
N-1a	0.90	0.93	--
O-3	1.53	1.40	--
Cat-Floc T Runs at 15 m/h (nominally)			
M-4a	1.22	1.05	--
M-4	0.86	0.98	--
M-4b	0.81	0.91	--
O-1b	1.29	1.25	--
O-1c	1.06	1.19	--
O-1	1.51	1.26	--
O-2	1.41	1.31	--

runs carried out with Cat-Floc T, the behavior was not so clear. For runs L, M, and N an older supply of Cat-Floc T was used; for these runs, the effect of flocculation was rather erratic: in some instances, flocculation promoted better effluent turbidity while in others, the opposite was achieved. However, for five runs carried out using a new supply of Cat-Floc T, flocculation consistently promoted higher turbidity removal.

Filter 2 processing flocculated water experienced more frequent breakthrough than Filter 3. In many instances terminal head loss was reached by Filter 3 before breakthrough started in that filter.

A filter run is normally terminated either when it reaches maximum available head loss, or when the filtrate turbidity reaches some maximum permissible level. For this research, the run time to the onset of breakthrough or to terminal head loss has been summarized in Table 37.

Table 37 reveals that when alum was used at naturally occurring pH, the effective run length was shorter with flocculation (Filter 2) because of earlier breakthrough. When alum was used with acid for pH control, the same trend was evident unless breakthrough did not occur. In the latter case, flocculation lengthened the effective filter run by reducing the rate of head loss development.

Similarly, when Cat-Floc T was used as primary coagulant, breakthrough did not occur; therefore, longer effective filter runs were achieved with Filter 2 than with Filter 3 due to the lower head loss accumulation in the former. This was true at both filtration rates.

The use of alum, without pH control, resulted in better average turbidity prior to breakthrough compared with the other two treatment schemes, but this benefit was offset by the short runs resulting from early breakthrough.

All of the chemical pretreatments used with or without flocculation produced filtered water well below 1 NTU for the remainder of the run after the first hour and prior to terminal breakthrough. However, when using either alum and acid or cationic polymer, 1 NTU average turbidity during the first hour was not achieved in many runs.

Particle Count

The results obtained for particle count in the range of 7-12 μm and 1-60 μm throughout the 16 formal runs are shown in Tables 38 and 39. In the same manner as before, each run was fragmented within three separate periods for the purpose of analysis and comparison of results, improvement period, remainder of the run prior to breakthrough, and breakthrough period. The improvement period was defined as the first hour of the filtration run. It was evident in many instances that improvement continued to occur after the first hour, especially for those runs carried out using Cat-Floc T.

The data in Tables 38 and 39 indicate similar results for both particle size ranges, with particle reductions in the 1 to 2 log range. In only one run in Table 39 was the reduction less than 1 log (Run O-1), and in many

TABLE 37. SUMMARY OF EFFECTIVE FILTRATION RUN LENGTH FOR FILTER 2, TREATING FLOCCULATED WATER AND FILTER 3, UNFLOCCULATED WATER

	Run Time (hours)*		Head Loss (cm)†	
Run Number	F2	F3	F2	F3
Alum Runs at 7.5 m/h (nominally)				
L-1	22 BT	29 BT	111	130
N-2a	38 BT	42 BT	102	162
N-2	27 BT	32 BT	87	127
N-2b	38 BT	42 BT	102	162
$\bar{x}$ =	31	36		
Alum Runs at 15 m/h (nominally)				
M-1c	6 BT	13	99	190
M-2	14.5 BT	15	129	190
M-2a	16	11	170	190
N-3a	13 BT	16 BT	142	175
N-3	7 BT	9	119	190
N-4	4.5 BT	9 BT	104	135
N-4a	24 BT	24 BT	135	147
N-4b	29 BT	36	168	190
$\bar{x}$ =	14	16		
Alum + Acid Runs at 7.5 m/h (nominally)				
L-2	14.5 BT	20 BT	142	175
M-1	49	40	190	190
P-1	34 BT	33 BT	109	105
$\bar{x}$ =	31	30		
Alum + Acid Runs at 15.0 m/h (nominally)				
M-3a	13 BT	15	127	190
M-3	14.5 BT	13	142	190
M-3b	10 BT	10	132	190
P-2	10 BT	11 BT	147	162
P-2a	18	15	190	190
$\bar{x}$ =	15	14		
Cat-Floc T Runs at 7.5 m/h (nominally)				
L-3	93	78.5	190	190
O-3	55	53	190	190
$\bar{x}$ =	74	66		

TABLE 37. CONTINUED

	Run Time (hours)*		Head Loss (cm)[†]	
Run Number	F2	F3	F2	F3
Cat-Floc T Runs at 15.0 m/h (nominally)				
M-4a	23.5	23.5	152	162
M-4	37.7	27	190	190
M-4b	40	35.5	190	190
O-1a	39	37	165	190
O-1b	36	32.5	190	190
O-1c	32.5	27	190	190
O-1	19	15	190	190
O-2	17.5	13.75	190	190
$\bar{x}$ =	29	25		

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[*] "To breakthrough" signified by BT or run termination of no breakthrough occurred.

[†] At breakthrough or run termination if no breakthrough occurred.

TABLE 38. SUMMARY OF AVERAGE 7-12 μm PARTICLE COUNT VALUES FOR THE FORMAL FILTRATION RUNS. VALUES IN No./mL

		Filter 2, with Flocculation			Filter 3, Unflocculated		
Run No.	Influent	First Hour	Remainder	Break-through	First Hour	Remainder	Break-through
Alum Runs at 7.5 m/h (nominally)							
L-1	1613(1)	190(3)*	149(2)	437(4)	166(3)	231(3)	346(3)
N-2	--	64(2)	56(6)	177(1)	62(2)	56(6)	134(1)
Alum Runs at 15. m/h (nominally)							
M-2	1014(3)	118(3)	73(5)	137(2)	98(3)	77(6)	--
N-3	1282(2)	63(3)	51(3)	139(1)	61(3)	64(4)	--
N-4	4810(1)	165(3)	107(5)	220(1)	83(3)	80(5)	104(1)
Alum + Acid Runs at 7.5 m/h (nominally)							
L-2	1650(2)	460(3)	260(2)	439(2)	543(3)	231(2)	165(2)
M-1	612(1)	72(2)	56(5)	--	77(2)	59(5)	--
P-1	974(2)	16(2)	36(8)	359(1)	19(2)	25(8)	100(1)
Alum + Acid Runs at 15 m/h (nominally)							
M-3	420(3)	100(2)	66(8)	117(1)	110(2)	54(8)	--
P-2	335(1)	93(2)	78(5)	119(2)	120(2)	84(5)	92(2)
Cat-Floc T Runs at 7.5 m/h (nominally)							
L-3	1002(2)	120(2)	67(6)	--	115(2)	74(5)	--
O-3	2915(3)	23(2)	37(7)	--	29(2)	32(7)	--
Cat-Floc T Runs at 15 m/h (nominally)							
M-4	514(3)	55(2)	42(9)	--	52(2)	41(7)	--
O-1	919(1)	29(2)	25(5)	--	29(2)	27(5)	--
O-2	3831(1)	36(2)	25(8)	--	34(2)	23(8)	--

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*Number in parenthesis following each particle count was the number of samples taken and averaged for the period.

TABLE 39. SUMMARY OF 1-60 μm PARTICLE COUNT VALUES FOR THE FORMAL FILTRATION RUNS. AVERAGES IN No./mL

		Filter 2, with Flocculation			Filter 3, Unflocculated		
Run No.	Influent	First Hour	Remainder	Breakthrough	First Hour	Remainder	Breakthrough
Alum Runs at 7.5 m/h (nominally)							
L-1	39,100(1)	2576(3)*	1145(2)	2844(4)	3934(3)	1603(2)	2827(4)
N-2	--	2746(2)	2623(6)	3326(1)	3987(2)	2943(6)	2733(1)
Alum Runs at 15 m/h (nominally)							
M-2	30,046(3)	2971(3)	1003(5)	1027(2)	2304(3)	1337(5)	--
N-3	32,493(2)	1367(3)	1206(3)	1987(1)	2190(3)	1613(4)	--
N-4	87,617(1)	2537(3)	2266(5)	2790(1)	3134(3)	1853(5)	1846(1)
Alum + Acid Runs at 7.5 m/h (nominally)							
L-2	36,320(2)	3144(3)	1312(2)	2276(2)	4363(3)	1330(2)	908(2)
M-1	19,316(1)	1578(2)	624(5)	--	2193(2)	851(5)	--
P-1	22,526(2)	535(2)	676(8)	6648(1)	526(2)	606(8)	2069(1)
Alum + Acid Runs at 15 m/h (nominally)							
M-3	22,072(3)	1321(2)	535(8)	1017(1)	2714(2)	935(8)	--
P-2	16,780(1)	1174(2)	1528(5)	2230(2)	2134(2)	1697(5)	1067(2)
Cat-Floc T Runs at 7.5 m/h (nominally)							
L-3	33,800(2)	7970(2)	1209(6)	--	5884(2)	1142(5)	--
O-3	59,398(3)	2430(2)	1252(7)	--	3537(2)	1039(7)	--
Cat-Floc T Runs at 15 m/h (nominally)							
M-4	17,168(3)	3223(2)	1717(9)	--	2658(2)	1383(7)	--
O-1	17,651(1)	1566(2)	2278(5)	--	3419(2)	3273(5)	--
O-2	89,491(1)	1242(2)	559(8)	--	1636(2)	675(8)	--

*Number in parenthesis following each particle count was the number of samples taken and averaged for the period.

cases, particle count reductions approached 2 log (99%). The particle numbers were generally higher the first hour of the run than the remainder of the run (24 of 30 cases in Table 38, 27 out of 30 cases in Table 39).

Again, the comparisons between Filter 2 and Filter 3 are the most important to the objectives of this Phase III. To facilitate those comparisons, the ratios of Filter 3 to Filter 2 have been compiled in Tables 40 and 41. These data support the similar observations presented earlier for turbidity. Flocculation generally improved the performance during the first hour of the run and during the remainder of the run up to breakthrough, i.e., (F3/F2)>1. This was especially noticeable in the 1-60 µm data for the alum runs, with or without pH adjustment. It is less evident for the Cat-Floc T runs in either size range.

The fact that flocculation also caused greater breakthrough is evidenced by ratios of (F3/F2) < 1 during the breakthrough period in both size ranges.

Head Loss

Table 42 summarizes the results on head loss analysis. The time to reach a selected terminal head loss or the time elapsed to reach a common head loss for runs that ended due to breakthrough appear in that summary. Table 42 also includes the ratio of the times to reach a common head loss for the two filters (F2/F3). It is evident from Table 42 that flocculation always reduces the rate of head loss development (Ratio F2/F3 > 1), usually substantially. The run time to a common head loss at 7.5 m/h filtration rate was from 10% to 26% longer when using flocculation; and at 15 m/h it was 27% to 46% longer. It must be remembered, however, that many of the alum runs had shorter effective run times due to early breakthrough, especially if flocculation was provided. In those cases, the lower head loss with flocculation was of no benefit.

Chlorophyll-a

Five samples were collected throughout Phase III to determine the ability of both filters to remove chlorophyll. Samples were always collected after the initial improvement period and before the terminal breakthrough period as observed on the turbidity vs time record. Table 43 summarizes the observations.

As the number of samples analyzed was limited to 5, such a small number prevents valid comparisons on the effectiveness of direct filtration with volume flocculation vs in-line filtration.

The chlorophyll levels were very low in all of these samples. At these 2 low levels, even though a sample of 0.5 or 1 L was filtered in the field to collect the algae, the resulting absorbence in the spectrophotometric chlorophyll analysis was very low, impairing the precision of the analysis. This may explain the inconsistent and rather low percentage removals shown in Table 43. The percent reductions in chlorophyll-a are substantially lower than the reductions in turbidity or particle count reported earlier.

TABLE 40. SUMMARY OF RATIO OF AVERAGE 7-12 μm PARTICLE COUNT RESULTS IN THE EFFLUENT FOR FILTER 3 RECEIVING UNFLOCCULATED WATER/FILTER 2 RECEIVING FLOCCULATED WATER

	Ratio of Average 7-12 μm Particle Count (F3/F2)		
Run Number	First Hour	Remainder	Breakthrough
Alum Runs at 7.5 m/h (nominally)			
L-1	0.88	1.55	0.79
N-2	0.96	1.00	0.76
Alum Runs at 15 m/h (nominally)			
M-2	0.83	1.06	--
N-3	0.97	1.25	--
N-4	0.50	0.75	0.47
Alum + Acid Runs at 7.5 m/h (nominally)			
L-2	1.18	0.89	0.38
M-1	1.07	1.05	--
P-1	1.21	0.71	0.28
Alum + Acid Runs at 15 m/h (nominally)			
M-3	1.10	0.82	--
P-2	1.07	1.07	0.78
Cat-Floc T Runs at 7.5 m/h (nominally)			
L-3	0.96	1.10	--
O-3	1.29	0.85	--
Cat-Floc T Runs at 15 m/h (nominally)			
M-4	0.94	0.98	--
O-1	0.97	1.05	--
O-2	0.95	0.92	--

TABLE 41. SUMMARY OF RATIO OF AVERAGE 1-60 μm PARTICLE COUNT RESULTS IN THE EFFLUENT FOR FILTER 3 RECEIVING UNFLOCCULATED WATER/ FILTER 2 RECEIVING FLOCCULATED WATER

	Ratio of Average 1-60 μm Particle Count (F3/F2)		
Run Number	First Hour	Remainder	Breakthrough
Alum Runs at 7.5 m/h (nominally)			
L-1	1.53	1.40	0.99
N-2	1.45	1.12	0.82
Alum Runs at 15.0 m/h (nominally)			
M-2	0.78	1.33	--
N-3	1.60	1.34	--
N-4	1.24	0.82	0.81
Alum + Acid Runs at 7.5 m/h (nominally)			
L-2	1.39	1.01	0.40
M-1	1.39	1.36	--
P-1	0.98	0.90	0.31
Alum + Acid Runs at 15.0 m/h (nominally)			
M-3	2.05	1.75	--
P-2	1.82	1.11	0.48
Cat-Floc T Runs at 7.5 m/h (nominally)			
L-3	0.74	0.94	--
O-3	1.46	0.83	--
Cat-Floc T Runs at 15.0 m/h (nominally)			
M-4	0.82	0.81	--
O-1	2.18	1.44	--
O-2	1.32	1.21	--

TABLE 42. SUMMARY OF FILTER RUN TIME REQUIRED TO REACH A SELECTED HEAD LOSS FOR FILTER 2 WITH FLOCCULATION AND FILTER 3, UNFLOCCULATED

Run Number	Selected Head Loss (cm)	Time to Reach H-L F2 (h)	 F3 (h)	Ratio (F2/F3)
Alum Runs at 7.5 m/h (nominally)				
L-1	172	48.5	46.0	1.05
N-2	133	42.0	33.5	1.25
N-2b	119	47.5	31.75	1.49
				$\bar{x}$ = 1.26
Alum Runs at 15 m/h (nominally)				
M-1c	180	20.0	12.0	1.66
M-2	190	24.5	16.0	1.53
M-2a	173	16.0	11.0	1.45
N-3a	190	24.0	17.5	1.37
N-3	132	10.0	8.25	1.21
N-3b	160	16.5	12.0	1.37
N-4	178	22.5	16.0	1.40
N-4a	190	48.0	35.0	1.37
N-4b	190	35.5	27.0	1.31
				$\bar{x}$ = 1.41
Alum + Acid Runs at 7.5 m/h (nominally)				
L-2	165	46.5	36.5	1.27
M-1a	140	29.0	25.0	1.16
M-1	190	49.0	39.75	1.23
P-1	127	41.75	36.5	1.14
				$\bar{x}$ = 1.20
Alum + Acid Runs at 15 m/h (nominally)				
M-3a	180	24.5	15.5	1.37
M-3	190	21.5	13.0	1.65
M-3b	190	18.5	10.5	1.76
P-2	190	14.0	11.5	1.21
P-2a	190	14.25	11.0	1.29
				$\bar{x}$ = 1.46

TABLE 42. CONTINUED

Run Number	Selected Head Loss (cm)	Time to Reach H-L F2 (h)	Time to Reach H-L F3 (h)	Ratio (F2/F3)
Cat-Floc T Runs at 7.5 m/h (nominally)				
L-3	190	93.0	78.5	1.18
O-3	190	55.0	53.0	1.03
				$\bar{x}$ = 1.10
Cat-Floc T Runs at 15 m/h (nominally)				
M-4a	152	23.5	18.5	1.27
M-4	190	37.5	27.0	1.38
M-4b	190	40.0	35.5	1.14
O-1a	165	38.75	28.25	1.37
O-1b	190	35.75	32.5	1.10
O-1c	190	32.5	27.0	1.20
O-1	190	19.0	15.0	1.20
O-2	190	17.5	13.75	1.27
				$\bar{x}$ = 1.27

TABLE 43. CHLOROPHYLL-A RESULTS AND PERCENT REDUCTIONS IN PHASE III RUNS

		Effluent			
		Filter 2		Filter 2	
Run	Influent (mg/m^3)	mg/m^3	Percent Reduction	mg/m^3	Percent Reduction
M-2 (A)*	3.5	1.2	(66)	1.3	(63)
M-3 (A-A)†	2.5	0.8	(68)	0.6	(76)
O-1 (P)‡	3.2	1.1	(66)	1.8	(44)
O-2 (P)	2.4	1.4	(42)	1.3	(46)
P-1 (A-A)	5.1	3.7	(27)	3.4	(33)

*Alum was used as primary coagulant.

†Alum was used as primary coagulant; the pH was adjusted with sulfuric acid.

‡Cat-Floc T was used as primary coagulant.

SECTION 11

FILTRATION DESIGN CONSIDERATIONS

SPECIAL NEEDS OF SMALL SYSTEMS

The general goal of this research was to explore effective filtration methods for small systems, especially those treating high-quality surface waters that might at times carry the cysts of *Giardia lamblia*. Small systems are arbitrarily defined as those serving less than 1,000 people or treating less than about 100,000 gallons per day. One must be cognizant of the typical conditions encountered in small systems of this size that should influence design decisions. Some of the typical conditions are:

1. The operator may be marginally trained and is often lacking in background in basic sciences to facilitate improvement.

2. The turnover of operating personnel is frequent, further compounding the training problem.

3. The system will operate unattended with intermittent visits of the operator rather than continuous operational attendance.

4. Because of the small plant size, only a few filters, two or three at most, will be provided.

5. The system is sometimes designed to operate fewer than 24 hours per day, although this is not common where slow sand filters are employed.

As a result of these conditions, certain goals should be satisfied in design of the small filtration system. It should:

1. Be simple to understand and to operate.

2. Be resistant to adverse effects of on-off operation upon filtrate quality if continuous operation is not possible.

3. Be fail-safe, without high technology apparatus, in case the operator fails to do his job.

4. Use filters designed to minimize deleterious effects created by sudden plant flow changes imposed by the operator.

5. Use filter media open to the atmosphere to encourage frequent inspection by operating personnel.

In view of these goals, three filtration suggestions are presented below that are equally appropriate for slow or rapid granular bed filtration systems:

1. Gravity filters should be used. This is common for slow sand filters, but pressure filters are sometimes used for rapid filters. Some disease outbreaks, traceable to pressure filters, were documented in the literature section. Recommended Standards For Water Works (i.e., Ten States Standards, 1982) do not allow pressure filters for polluted surface waters or for direct filtration applications [66].

2. The filter effluent pipe should exit above the top of the filter media. This will prevent the development of negative head in the filter media and prevent accidental dewatering of the filter in the event of cessation of the influent flow to the filter.

3. Influent flow splitting should be used because it provides many advantages [15,88]. During on-off operation, the filtration rate increases or decreases slowly and smoothly without the need for automatic equipment. Similarly, if plant flow changes are made by the operator, the filtration rate changes occur slowly and smoothly, thus minimizing detrimental effects on the filtrate quality. This system also eliminates the need for rate of flow controllers, head loss gauges, and individual flow meters on each filter. It is the simplest system of control for the operator to understand. When the water level reaches an upper limit at maximum available head loss, the operator must take action to clean the filter or overflow will occur. An overflow to waste should be provided for such emergencies.

THE CHOICE BETWEEN SLOW AND RAPID FILTRATION

Slow Sand Filtration

From the foregoing conditions and goals, and from the superior performance of the slow sand filter demonstrated in this research, it would seem obvious that the slow sand filter is the system of choice for high-quality surface waters that might contain the cysts of Giardia lamblia, but are consistently low in algae, color, turbidity, taste, and odor.

The advantage of the slow sand filter is its simplicity and its immunity to operator abuse. It requires no chemical coagulation prior to pretreatment, so no knowledge of chemistry or chemical coagulation is required. The cycle length between cleanings can be several weeks or months. The daily surveillance of the plant disinfection equipment, which will be required in any case, will afford the opportunity to observe that the slow sand filter is operating normally. The operator can observe the head loss development during such visits, and perform any necessary sampling and monitoring tests of the filtrate quality.

When the head loss reaches an upper limit, the filter must be drained and scraped. This is usually done manually and would not be an excessively burdensome task for very small systems.

Therefore, if land can be obtained and if the raw water quality is appropriate, the slow sand filter should be given serious consideration. The principal question is what raw waters are appropriate for the slow sand filter.

Appropriate Raw Waters for Slow Sand Filtration

Ten States Standards limit the slow sand filter to waters with "maximum turbidities of 50 units and maximum color of 30 units; such turbidity must not be attributable to colloidal clay. Raw water quality data must include examinations for algae" [66]. These suggestions are too vague to be useful. From the experience of the research reported here, a water with average turbidity of 50 NTU would not be appropriate, but a short-term peak turbidity of 50 might be tolerated if it was not sustained more than a few hours. Since the slow sand filter is not very effective in color removal, a raw water of 30 color units may not meet the recommended limit of 15 color units after filtration. This limit is recommended as a secondary standard under the U.S. Safe Drinking Water Act. The manner of using the algae information in design decisions is not stipulated by Ten States Standards.

From a review of the slow sand filtration experiments presented here, filter runs greater than 30 days occurred when algal populations were low as evidenced by chlorophyll-a measurements less than 5 mg/m^3, accompanied by average turbidities under 5 NTU. Short-term peak turbidities between 10 and 30 NTU occurred for a few hours during these runs and were handled without difficulty. These peaks were usually caused by strong winds that caused bank erosion, contributing clay and silt particles to the raw water intake.

In contrast, some runs during algal blooms were as short as 9 days, in spite of the fact that average turbidity was still near 5 NTU. Thus, turbidity alone was not an adequate predictor of run length for slow sand filtration when algae were also present. Evaluation of the extent of the algal problem by direct counting of algae or by a surrogate measure such as chlorophyll-a should be conducted for at least a full year to observe the full seasonal range of the problem. From the data reported here, waters with an average of 5 mg/m^3 of chlorophyll-a along with concurrent average turbidity of 5 NTU would be appropriate raw waters for slow sand filtration. Of course, many mountain raw waters are well below these limits and would be appropriate for slow sand filtration.

Appropriate Raw Waters For Direct, Rapid Filtration

For somewhat poorer waters, for larger systems, or if land is unavailable for slow sand filters, direct, rapid filtration would be one alternative to be considered. It must be viewed, however, as a higher level of technology requiring substantially more operator training. Some knowledge of chemistry and chemical coagulation are essential for the operator; therefore the direct, rapid filtration alternative should be considered along with other alternatives such as precoat filtration using diatomaceous earth.

Before selecting direct, rapid filtration, the question of which raw water characteristics are appropriate for direct, rapid filtration must be answered.

The literature on this question has been presented in the literature section of this report. Before attempting to answer this question, one must again pay attention to the special requirements of small systems, particularly with regard to minimum acceptable run length. Large plants with operators present 24 hours per day can tolerate shorter filter runs than small plants with intermittent operator attendance. Plant production is not impaired if the unit filter run volume (UFRV, i.e., volume produced per unit area per run) does not fall below about 200 m^3/m^2/run (5000 gal/ft^2/run). With typical filtration rates of 10 to 12.5 m/h (about 4-5 gpm/ft^2) in large plants today, this UFRV results in acceptable minimum run lengths of (200 m^3/m^2/run)/(12.5 m^3/m^2/h) = 16 h/run at 12.5 m/h (5 gpm/ft^2). Some large plants manage to survive periods with runs as short as 6 to 8 hours, although with some loss of production efficiency and a good deal of operational inconvenience.

However, for small plants attempting to operate with only a daily visit of the operator to the plant to backwash the filters, restock chemical feeders, and do the necessary sampling and analysis work, etc., filter runs of 24 hours would be minimum acceptable length. This would imply an upper filtration rate of 200/24 = 8.3 m/h (3.4 gpm/ft^2) if the UFRV is at the lower limit of 200 m^3/m^2/run.

A review of the results in Table 20 for Phases I and II of this research indicates that run lengths of 24 h or longer were generally achieved with filtration rates of 7.5 m/h (nominally) except during severe algal blooms. During the severe algal blooms of July 1982, with average raw turbidities up to 18 NTU, runs were as short as 8 hours and were terminated by breakthrough rather than reaching terminal head loss. The use of prechlorination was beneficial to the filtrate quality during such periods, but run lengths were still less than 24 h and were terminated by breakthrough.

The data from Phase III in Table 37 also indicates run lengths generally longer than 24 h at filtration rates of 7.5 m/h (nominally). The alum runs were frequently terminated by breakthrough rather than reaching terminal head loss, whereas the Cat-Floc T runs were never terminated by breakthrough and generally reached a terminal head loss of 1.9 m.

In all phases of the study, the Cat-Floc T runs were substantially longer than alum runs when treating similar raw waters in adjacent time periods, other factors being equal.

In an attempt to define upper limits of raw water quality appropriate for direct filtration, the data for the filter runs have been used to calculate the solids load applied per unit filter area per unit head loss increase ($Kg/m^2/m$). This is admittedly a simplistic approach that assumes a near-linear head loss development proportional to the solids applied or removed by the filter. Since percent removal was generally rather high in this direct filtration study, applied load was selected because it was nearly equal to solids removed and was simpler to calculate and use. In order to convert the average turbidity of the raw water to equivalent suspended solids concentration, the correlations previously presented in this study were used. They are repeated here for convenience:

For turbidity from 0-5 NTU

$$SS(mg/L) = 1.25 \cdot Turbidity(NTU) + 0.46$$

For turbidity from 5-17 NTU

$$SS(mg/L) = 1.83 \cdot Turbidity(NTU) - 1.11$$

The solids load values were calculated as follows:

$$\begin{array}{c}\text{Solids load}\\ (Kg/m^2/m\Delta H.L.)\end{array} = \frac{\text{avg raw } SS(Kg/m^3) \cdot \text{rate}(m/h) \cdot \text{run time}(h)}{\text{final head loss (m)} - \text{initial head loss (m)}}$$

The calculated values are summarized in Table 44 for Phases I and II of the study. The other conditions of the runs can be observed in Table 20.

T44

The solids load values in Table 44 are remarkable in their variability, even for a particular chemical treatment; thus, one must not place too much reliance on mean values. As stated in the report, the raw water was of good quality in the fall of 1981, but contained more algae and was more difficult to treat beginning with the 1982 snow melt and for the rest of 1982. The mean solids loads for these two periods and two principal chemical treatments are as follows:

	Solids Load ($Kg/m^2/m$)	
	Using Alum	Using Cat-Floc T
Fall '81	1.9	2.5
Snow melt thru Fall '82	1.1	1.8

The data taken during ice-cover were excluded in calculating the mean values because the very clear water during this period resulted in very long

TABLE 44. SOLIDS APPLIED PER UNIT AREA PER UNIT INCREASE IN HEAD LOSS DURING PHASES I AND II

Season	Run Number	Coagulant Used	Solids Load ($Kg/m^2/m$)	Termination Cause*
Fall 81	A-1	Alum + Acid	2.09	BT
	A-2	"	1.56	
	A-3	"	0.88	
	A-4	"	2.44	
	A-5	"	1.39	
	B-1	"	3.12	
	B-2	Cat-Floc T	3.38	
	B-3	"	1.65	
Winter 81-82 Ice Cover	B-4	"	0.86	
	B-7	"	1.10	
	B-8	"	1.43	
Snow melt '82	B-11	Alum + Acid	1.24	BT
	B-13	Cat-Floc T	1.47	
Spring '82 Ice Gone	B-15	Alum + Acid	1.20	
	C-1	"	0.66	
	C-3	Alum	0.94	BT
	C-4a	Alum + Acid	1.30	BT
Summer '82	E-1	Alum + Acid	1.03	BT
	G-1	Cat-Floc T	2.20	
	H-1	Alum + Cl_2	1.78	BT
	H-2f	Alum + Acid + Cl_2	1.81	BT
	I-4	Alum + Cl_2	0.27	
	I-5	"	1.54	
	I-6c	Cat-Floc T	2.02	
	I-6e	"	1.97	
Fall '82	J-1	Alum	1.64	BT
	J-2	"	0.44	
	J-3	Cat-Floc T	1.59	
	J-4	"	1.57	
	J-6	Alum	0.82	
	J-7	"	0.92	
	J-8	Cat-Floc T	1.59	
	J-9	"	1.99	

---

*BT means terminated by breakthrough before reaching terminal head loss. All other runs did not exhibit terminal breakthrough.

filter runs, even though the solids load values per unit head loss increase were low during this period. Thus, these values were not particularly useful to short filter run problems of interest here.

The benefit of cationic polymer is evident in these mean values. It can also be observed in Table 44 by comparing adjacent runs in close succession with the two chemicals (for example, compare runs J-6 and J-7 with Runs J-8 and J-9).

Using the mean values given above, one can back-calculate the levels of suspended solids permissible to achieve a desired run length of 24 h at a filtration rate of 7.5 m/h (3 gpm/ft^2) with a head loss increase of 2 m. From the SS value, the equivalent turbidity value can be calculated from the regression equations given previously. The results are as follows:

	Permissible Raw Water Quality	
	Average Turbitidy (NTU)	Average SS (mg/L)
During low algae		
Using alum	12	21
Using cationic polymer	16	28
During moderate algae		
Using alum	7	12
Using cationic polymer	11	20

The above system of predicting run length can not be applied during periods when breakthrough necessitates run termination before reaching the full available head loss increase, 2 m in this case. The runs terminated by breakthrough are shown in Table 44, and were particularly common during heavy algal blooms in the summer of 1982.

The above permissible values generally are in harmony with some of the suggestions in the literature section. For example the AWWA committee limit of 15 NTU turbidity on a continuous basis [20] and McCormick and King's suggestion of 10 NTU [58] are quite similar. However, Culp's early suggestion of 200 turbidity units appears far beyond the reasonable range.

The above permissible values assume that color is not present in sufficient amount to require added coagulant dosage. If color is present, the values from the literature must be used as a guide. If the total alum dosage required for the color and turbidity removal is more than about 12 mg/L (as filter alum), the water should be considered a doubtful candidate for direct filtration. Pilot studies would be justified before designing the full-scale plant to determine the coagulant dosage required for color removal during the most critical season.

MISCELLANEOUS DESIGN CONSIDERATIONS

In the following pages, a few miscellaneous design considerations of special concern are discussed. It is not the intent of this section to present all details of filter design, which are available in textbooks [88] and design standards [66].

For Direct, Rapid Filtration

The following special aspects of direct, rapid filtration should also be considered in plant design.

Because the direct filtration plant has a short detention time of only a few minutes, there is less time to respond to changes in the raw water quality. Therefore, this process is most appropriate for waters of fairly stable quality or of quality that shifts gradually, such as in lakes or reservoirs. Provision should be made for continuous monitoring and recording of the raw and finished water turbidity, and for the fail-safe shutdown of the plant if the finished water shows a sudden deterioration, signalling a failure of coagulant feed or a sudden change in raw water.

If the plant is to operate unattended for a portion of each day, the chlorine content of the finished water should also be monitored continuously and automatically, with a fail-safe shutdown in the event of chlorine residual failure.

After automatic shutdown caused by failure of either turbidity or chlorine, manual start-up of the plant should be required to ensure that someone visits the plant and attempts to diagnose the cause of the shutdown and make a suitable remedial response. Automatic start-up should not be provided.

Direct, rapid filtration sometimes exhibits a lengthy initial improvement period; thus, the design should include provisions for convenient filtration to waste during the early period of each filter run. This could be manually controlled, or it could be automated to filter to waste for a preset time or until the filtrate reached a preset turbidity level. At this time, the filtrate should be automatically diverted to the treated water reservoir without causing a hydraulic disturbance to the filter.

Because of the short detention time in direct filtration, there is not adequate time for normal taste and odor control measures (e.g., powdered activated carbon treatment or chemical oxidation) to be effective. Thus, if taste and odor are expected to be a problem, a pretreatment basin is appropriate to provide the necessary detention time [20].

Because of the small number of filters in a small plant (2 or 3 filters) the removal of one filter from service for backwashing represents a substantial portion of the flow (33% to 50%). If that flow is diverted to the other filters during the backwash of a dirty filter, the other filters remaining in service would be subjected to a flow increase, (50% to 100%), causing a hydraulic shock and temporary deterioration of the filtrate. Therefore, for small plants, it is suggested that the influent to the filter be left open during the backwash, thus wasting the influent flow during the backwash period and avoiding the hydraulic shock to the other operating filter(s). This is especially important where Giardia lamblia cysts might be present.

The filter media for direct rapid filtration has been discussed in detail in the literature section of the report and this information will not be repeated here. If a coarse dual media is selected to handle high expected diatom populations, it must be recognized that a higher backwash rate is required to fluidize the coarse anthracite. The required rate is a function of the coarse grain size included in either the anthracite or sand layer, the density of the media, and the maximum expected temperature of the backwash water. Typical media designs and backwash rates are illustrated in Table 45 to emphasize the importance of this issue. It is evident that the selection of the coarse media should only be made when absolutely necessary because of the presence of filter clogging diatoms. The backwash rate includes a 30% safety factor above minimum fluidization velocity. It may be possible to succeed with a lower safety factor, but new plants should be designed with this capability. The backwash rate agrees well with the rate used in Phase II, which is reported in Table 9.

T45

Furthermore, to allow for potential backwashing difficulties created by the use of organic polymers in coagulation, filters should be equipped with auxiliary scour systems to assist the backwashing. These include either surface wash systems or air scour systems that have been properly designed according to conventional practice.

For Slow Sand Filtration

Only two special design precautions will be discussed for slow sand filtration dealing with the filter media and the need for housing the filter.

The Recommended Standards for Water Works [66] requires that the filter sand have an es between 0.3 and 0.45 mm and a uc not to exceed 2.5. The early literature cited would favor a size near this minimum or smaller. The 0.45 mm is too large and actually is up in the range of sand size for rapid filters.

However, if a fine size is selected near 0.3 mm es, the underdrain gravel must be designed to prevent passage of the fine sand through the supporting gravel. The Recommended Standards [66] suggest the same gravel design for the slow sand filter as for the rapid filter. This would not be appropriate if the sand is near 0.3 mm es. The gravel design used in this research (Table 3) would be a more appropriate design for this sand size. If the size of the underdrain openings were greater than 1/4 inch, one additional layer of gravel should be provided at the bottom with size range from 3/4 inch to 1.5 inch.

The Recommended Standards [66] suggest that the slow sand filter have a cover with adequate head room for access for scraping, etc. Traditionally, slow sand filters in warm climates have been built without a cover or housing of any sort. However, in severe climates as in the northern U.S. and in the

TABLE 45. TYPICAL DUAL MEDIA DESIGNS AND REQUIRED BACKWASH RATES

	Conventional Dual Media	Coarse Dual Media
Anthracite		
es (mm)	0.9-1.1	1.4-1.6
uc	≤ 1.5	≤ 1.5
d_{90} size (mm)	1.8-2.2	2.8-3.2
density (g/cm^3)	1.65	1.65
fluidization velocity* (mm/s)	11.5	18.8
backwash rate (mm/s)	15.0	24.4
(gpm/ft^2)	22	36
Sand of Compatible Size		
es (mm)	0.5-0.6	0.75-0.85
uc	≤ 1.5	≤ 1.5
d_{90} size (mm)	1.0-1.2	1.5-1.7
density (g/cm^3)	2.65	2.65
fluidization velocity* (mm/s)	11.0	17.6
backwash rate (mm/s)	14.3	22.9
(gpm/ft^2)	21	34

*For summer water temperatures of 20° C, and using the mean of d_{90} size range for the diameter, calculated with the Wen and Yu equation.

mountains, a cover should be provided. One can not always predict whether the winter run length will be long enough to carry through the freezing season. If it became necessary to scrape the filter in midwinter, the ice layer on an uncovered filter would end up on the sand and prevent the scraping operation. Therefore, in spite of the extra cost, the housed filter is strongly recommended where ice would develop on an uncovered filter.

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APPENDIX A: TURBIDITY VS. TIME, PHASE II

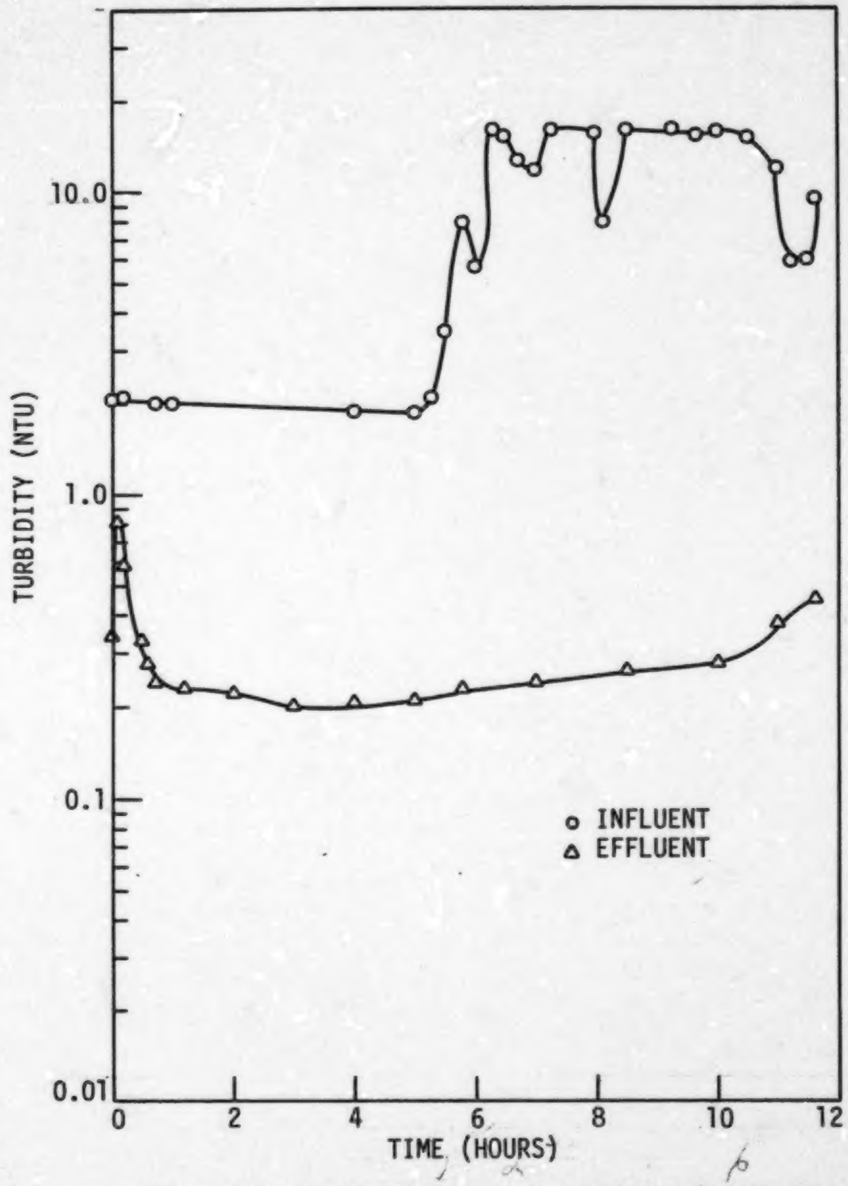
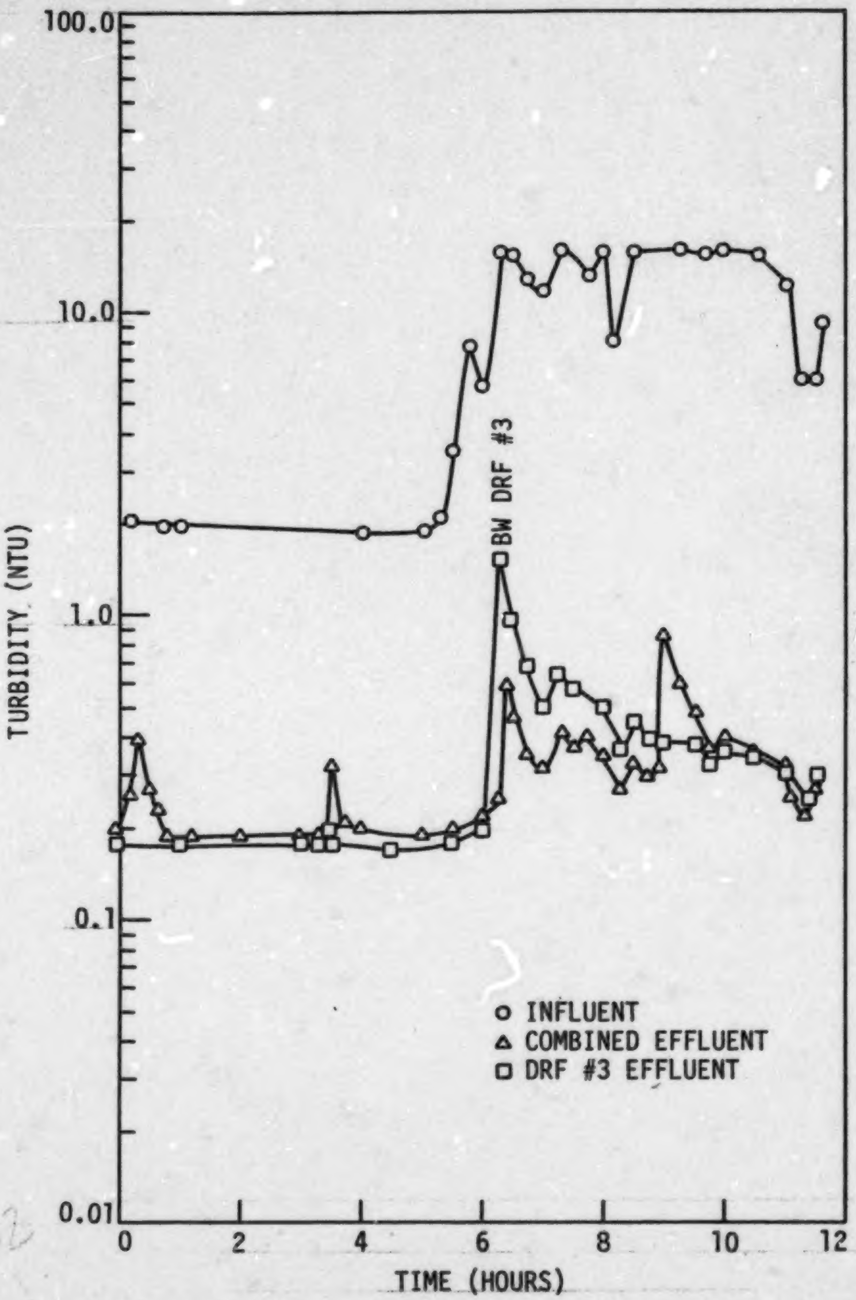


Fig. A-1. Turbidity for CRF during Run J-1 at 13.2 m/h (5.30 gpm/ft^2) using alum coagulant. Turbidities indicated as 16 could be greater than 16 NTU. The recorder was at full scale during this period. A single ratio turbidimeter reading of 16 NTU was obtained during this period.



A-2

Fig. A-2. Turbidity for DRF during Run J-1 at 13.2 m/h (5.30 gpm/ft^2) using alum coagulant.

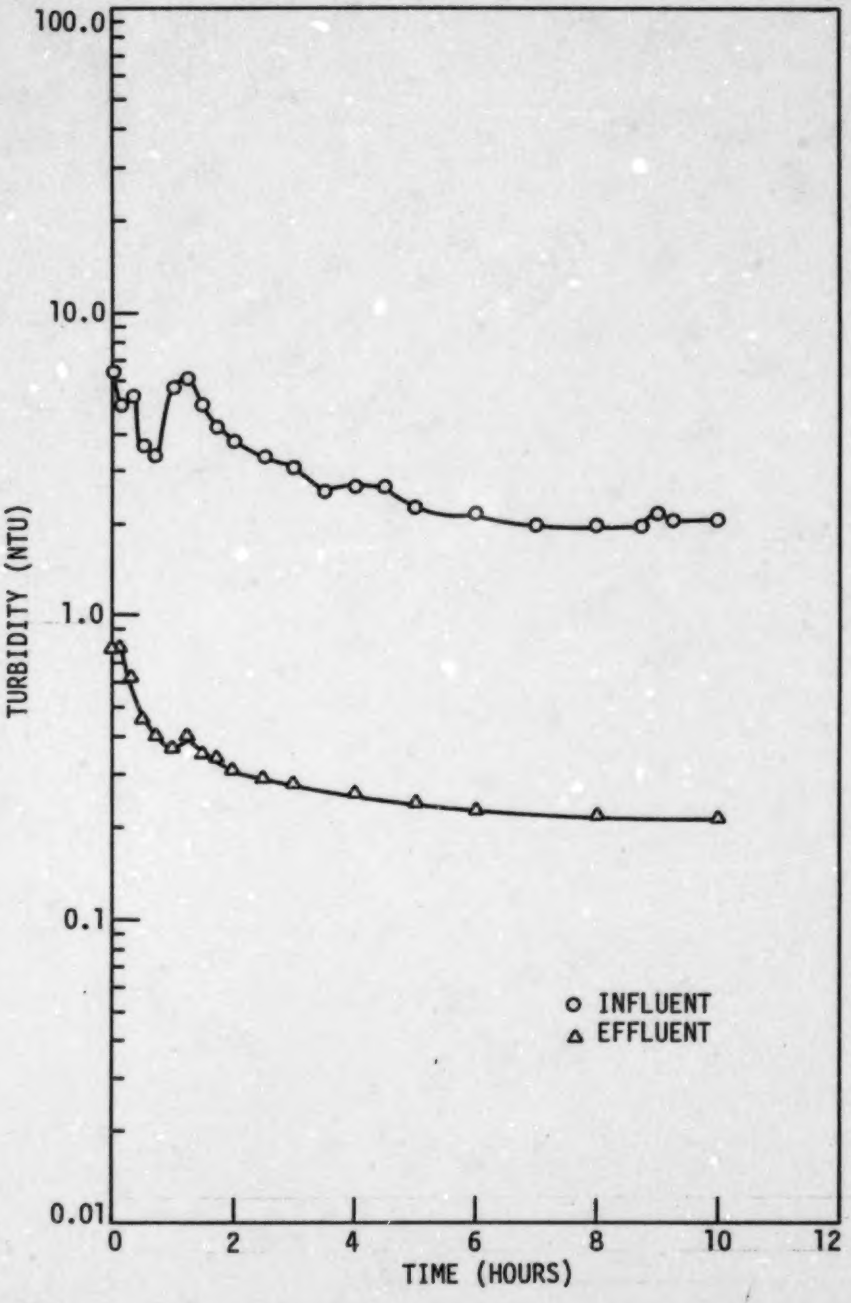


Fig. A-3. Turbidity for CRF during Run J-2 at 13.5 m/h (5.42 gpm/ft^2) using alum coagulant.

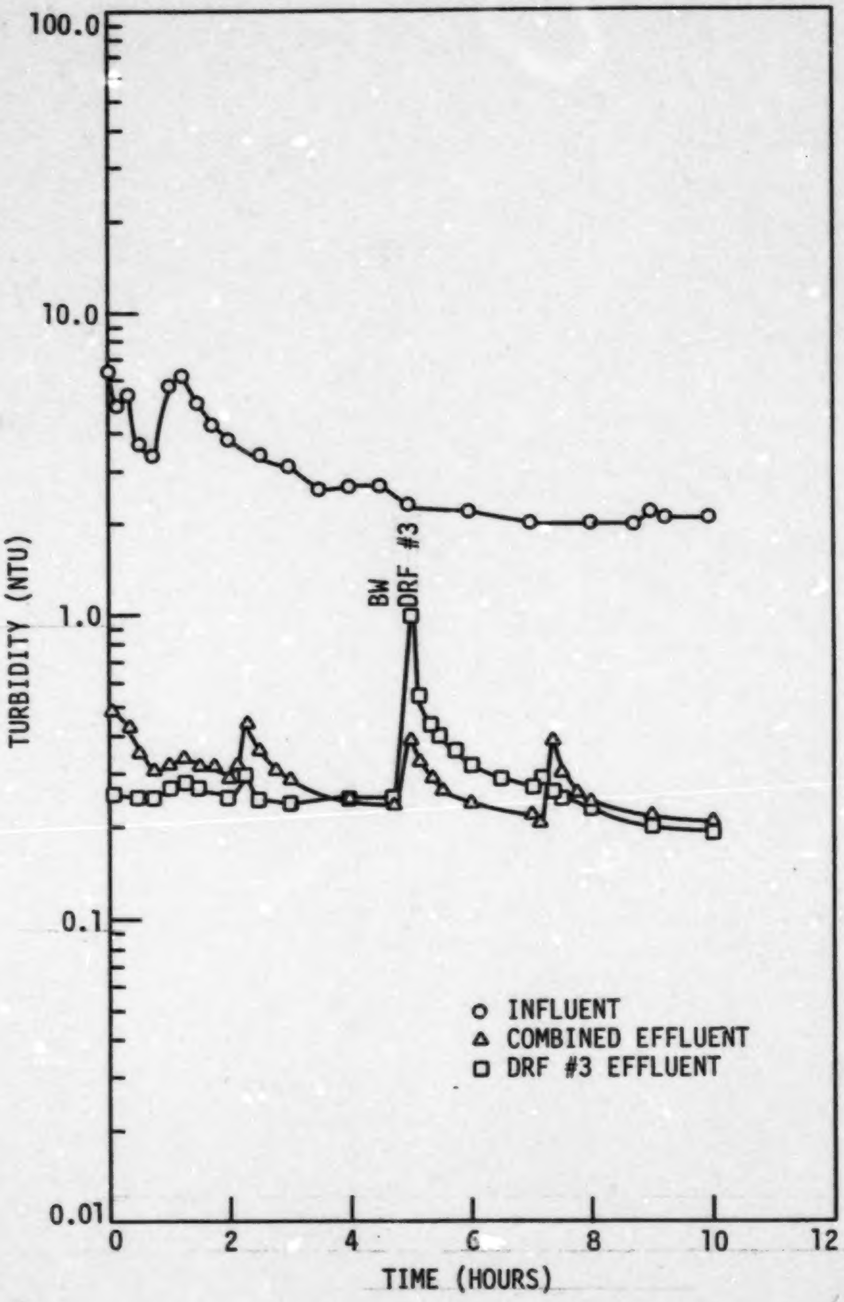


Fig. A-4 . Turbidity for DRF during Run J-2 at 13.5 m/h (5.42 gpm/ft^2) using alum coagulant.

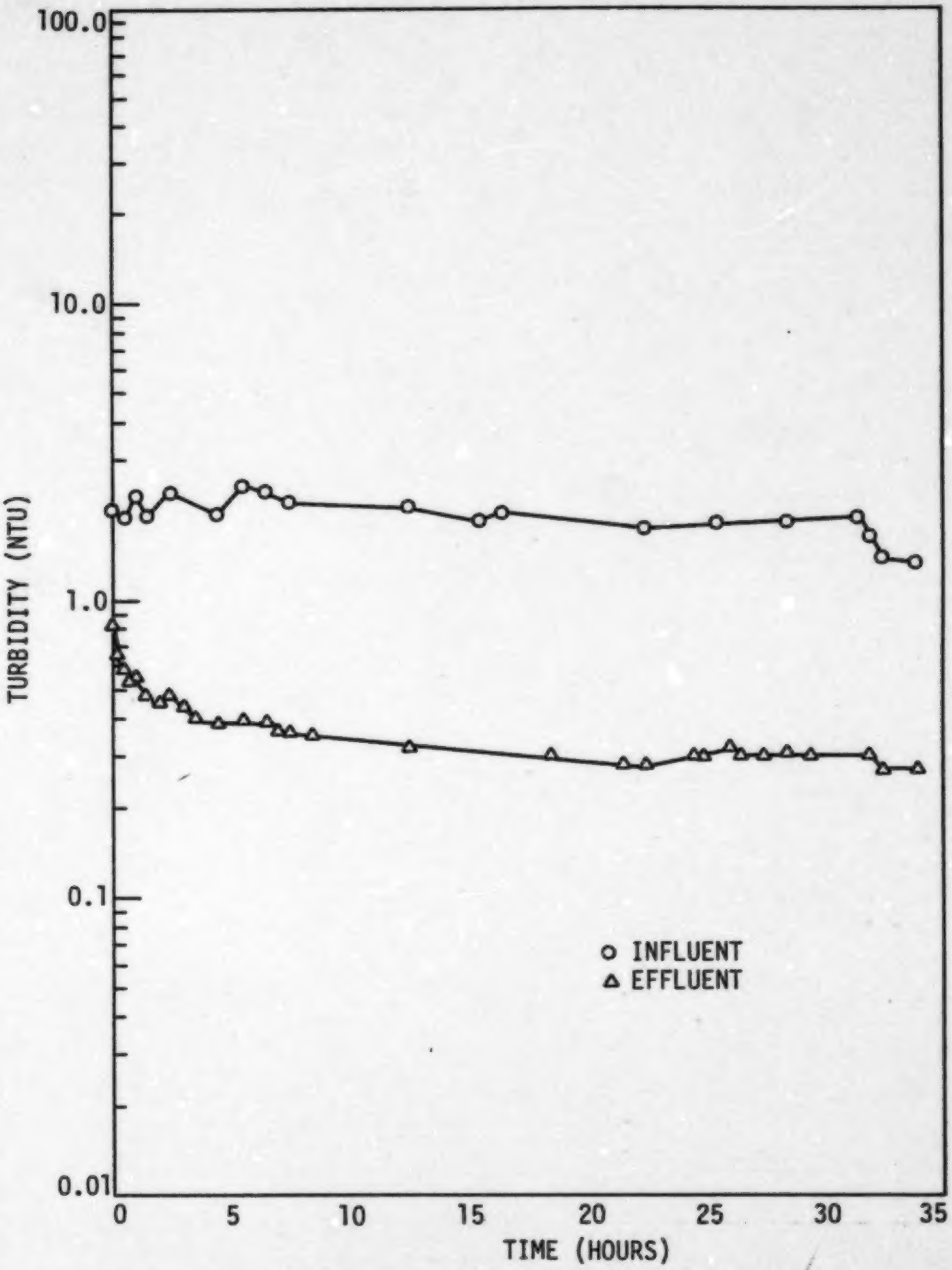


Fig. A-5. Turbidity for CRF during Run J-4 at 13.4 m/h (5.38 gpm/ft^2) using Cat-Floc T coagulant.

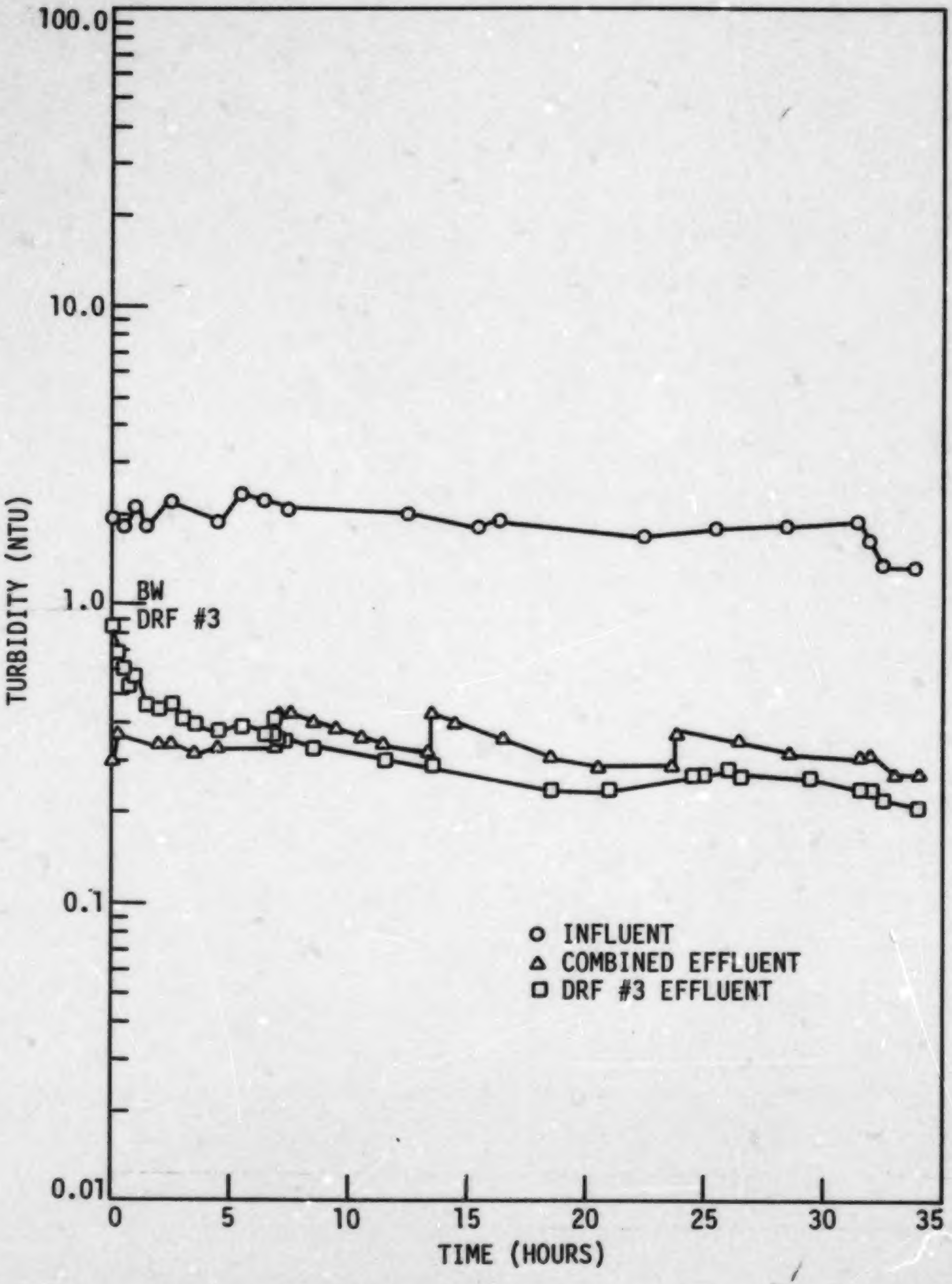


Fig. A-6. Turbidity for DRF during Run J-4 at 13.4 m/h (5.38 gpm/ft^2) using Cat-Floc T coagulant.

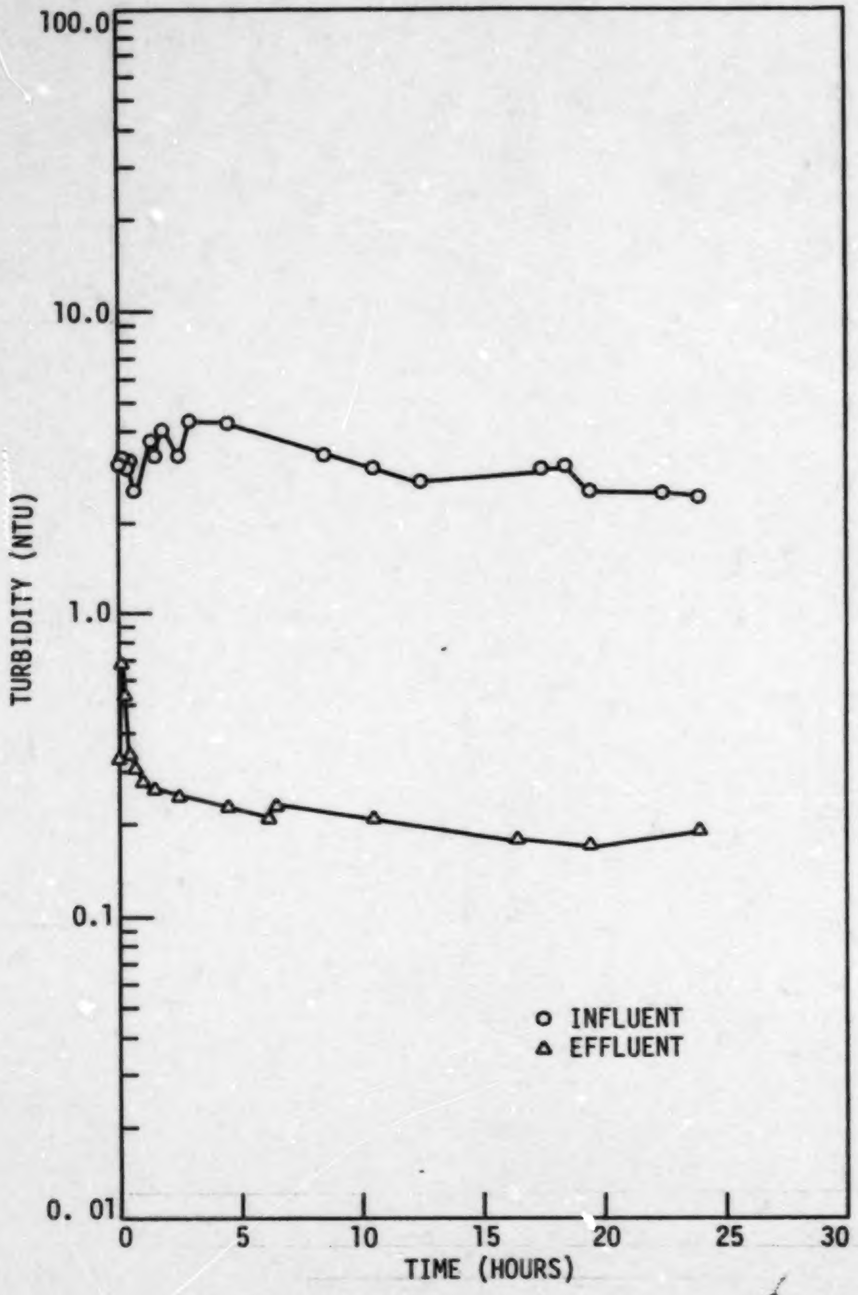


Fig. A-7. Turbidity for CRF during Run J-6 at 7.7 m/h (3.1 gpm/ft^2) using alum coagulant.

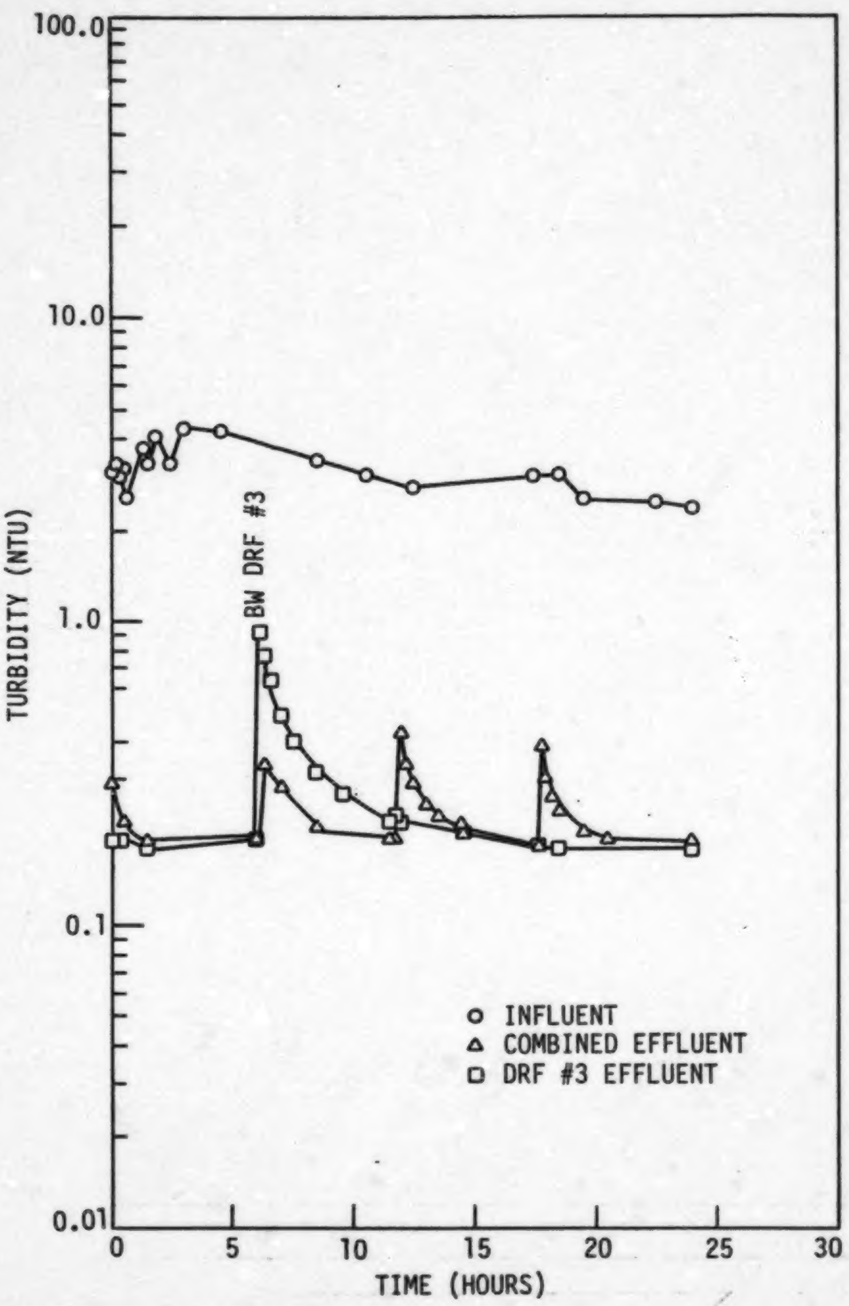


Fig. A-8. Turbidity for DRF during Run J-6 at 7.7 m/h (3.1 gpm/ft^2) using alum coagulant.

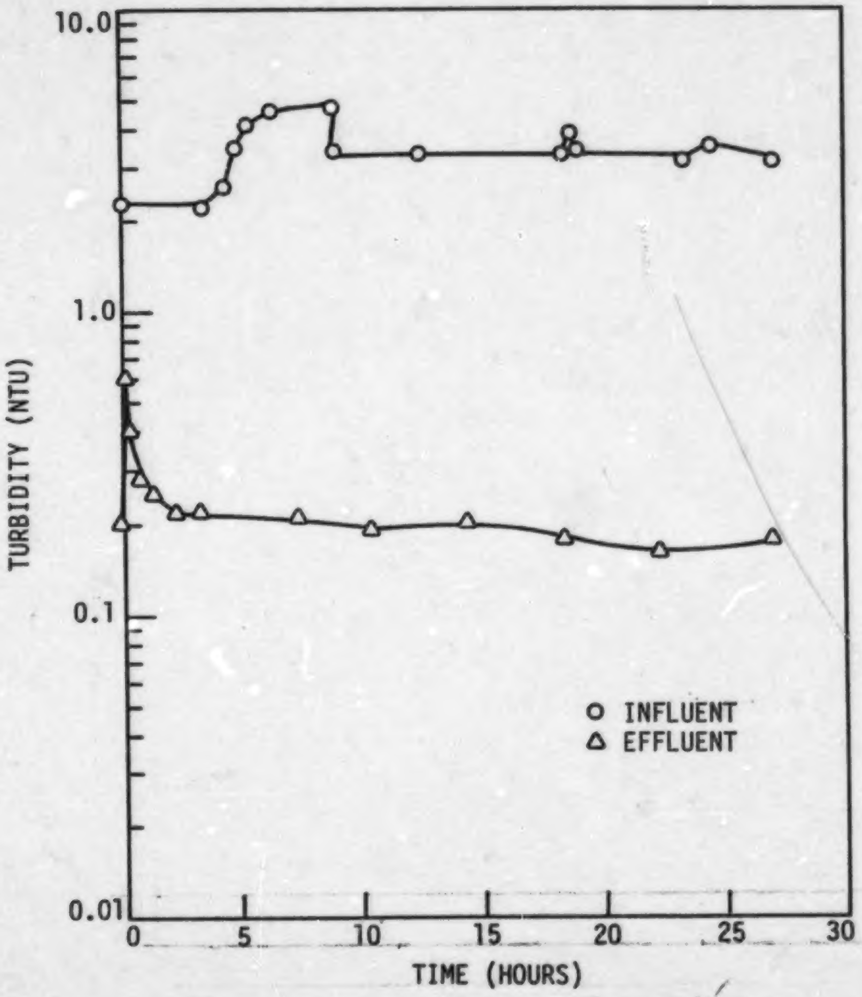


Fig. A-9. Turbidity for CRF during Run J-7 at 7.6 m/h (3.1 gpm/ft^2) using alum coagulant.

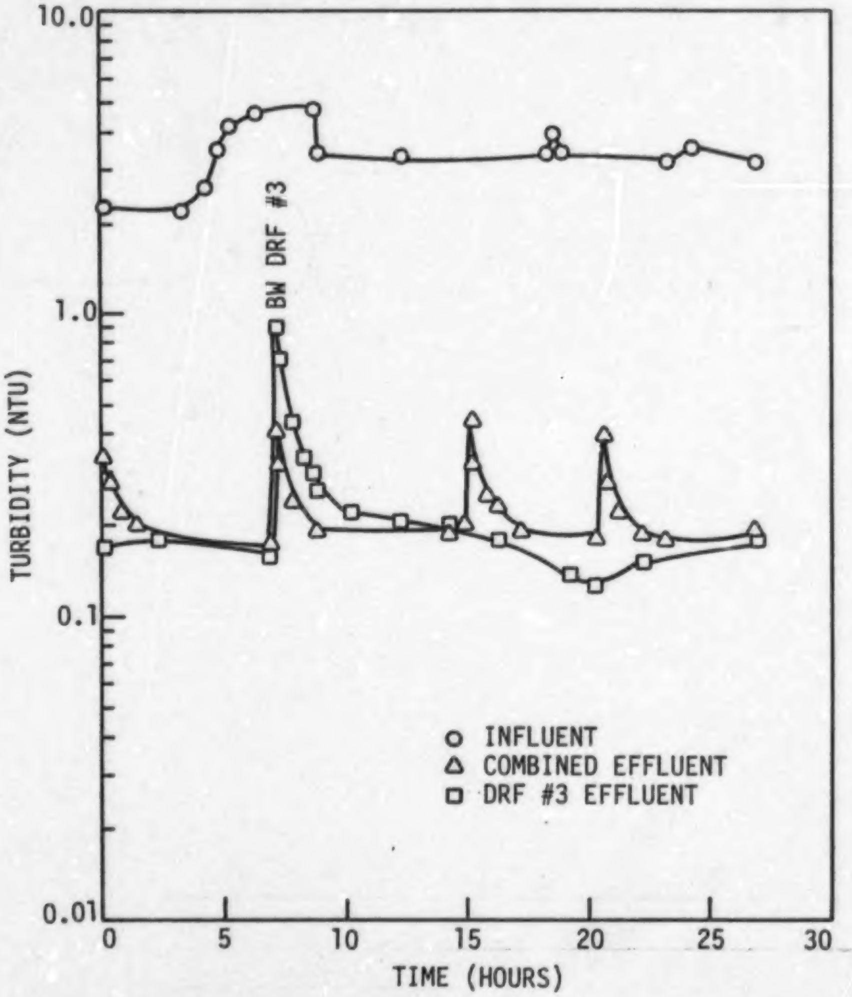


Fig. A-10. Turbidity for DRF during Run J-7 at 7.6 m/h (3.1 gpm/ft^2) using alum coagulant.

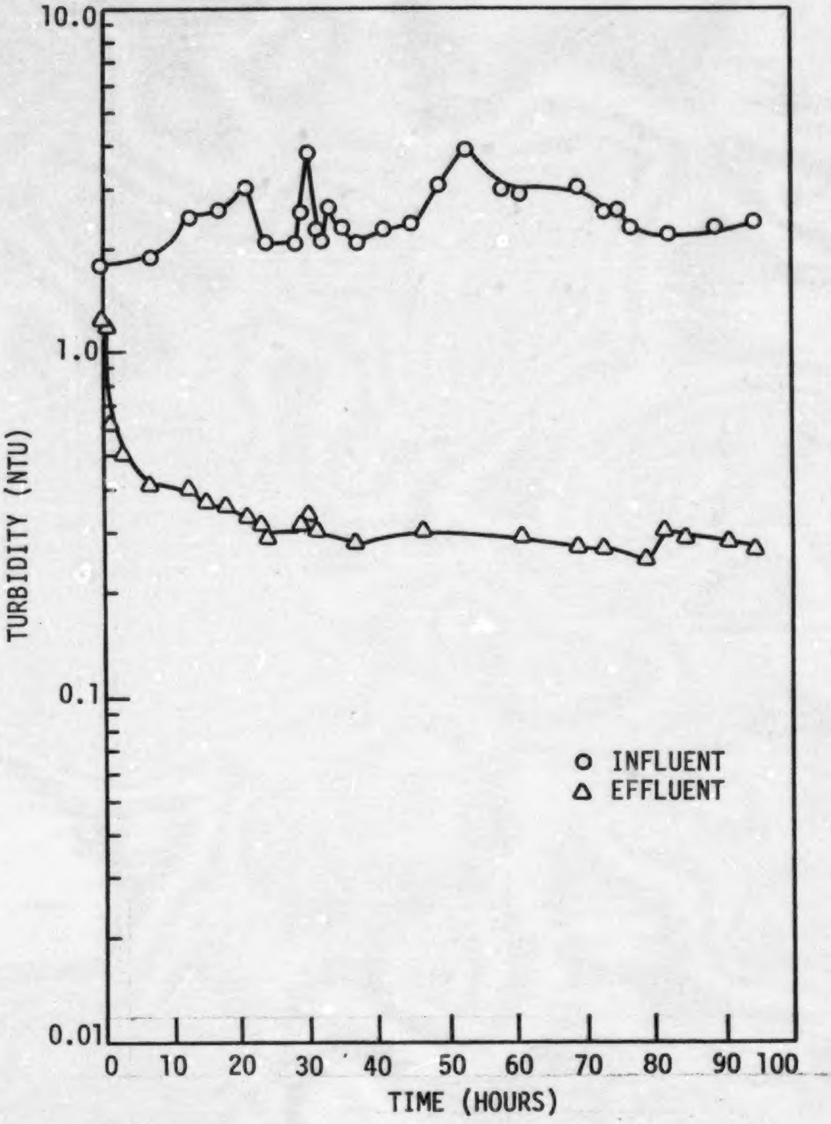


Fig. A-11. Turbidity for CRF during Run J-8 at 7.7 m/h (3.1 gpm/ft^2) using Cat-Floc T coagulant.

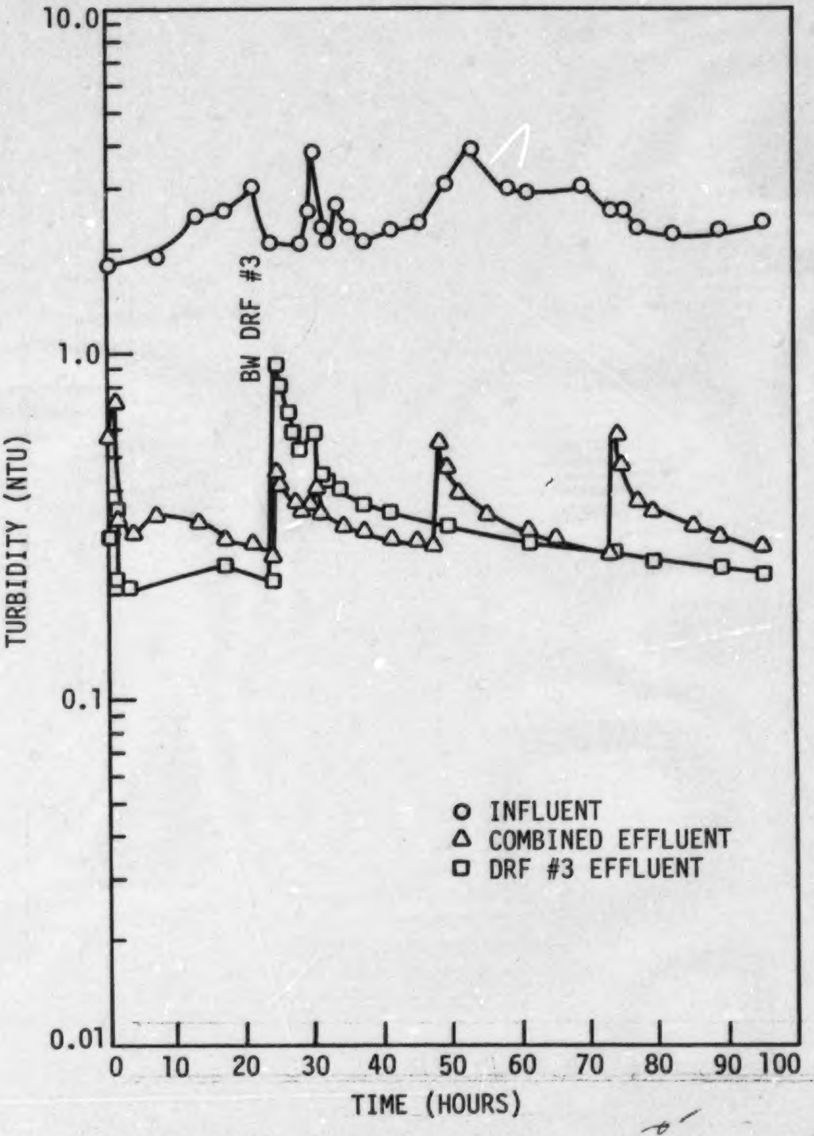


Fig. A-12. Turbidity for DRF during Run J-8 at 7.7 m/h (3.1 gpm/ft^2) using Cat-Floc T coagulant.

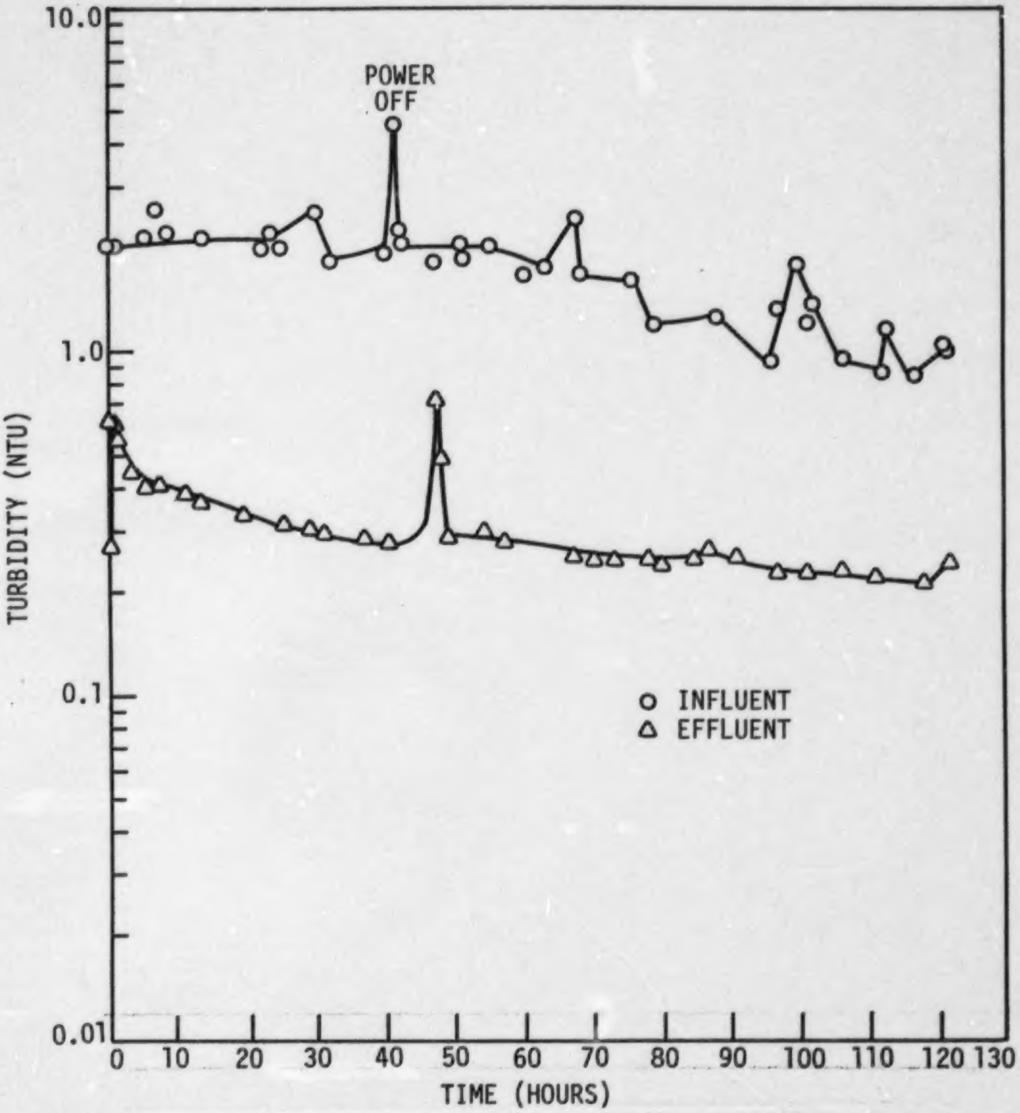


Fig. A-13. Turbidity for CRF during Run J-9 at 7.7 m/h (3.1 gpm/ft^2) using Cat-Floc T coagulant.

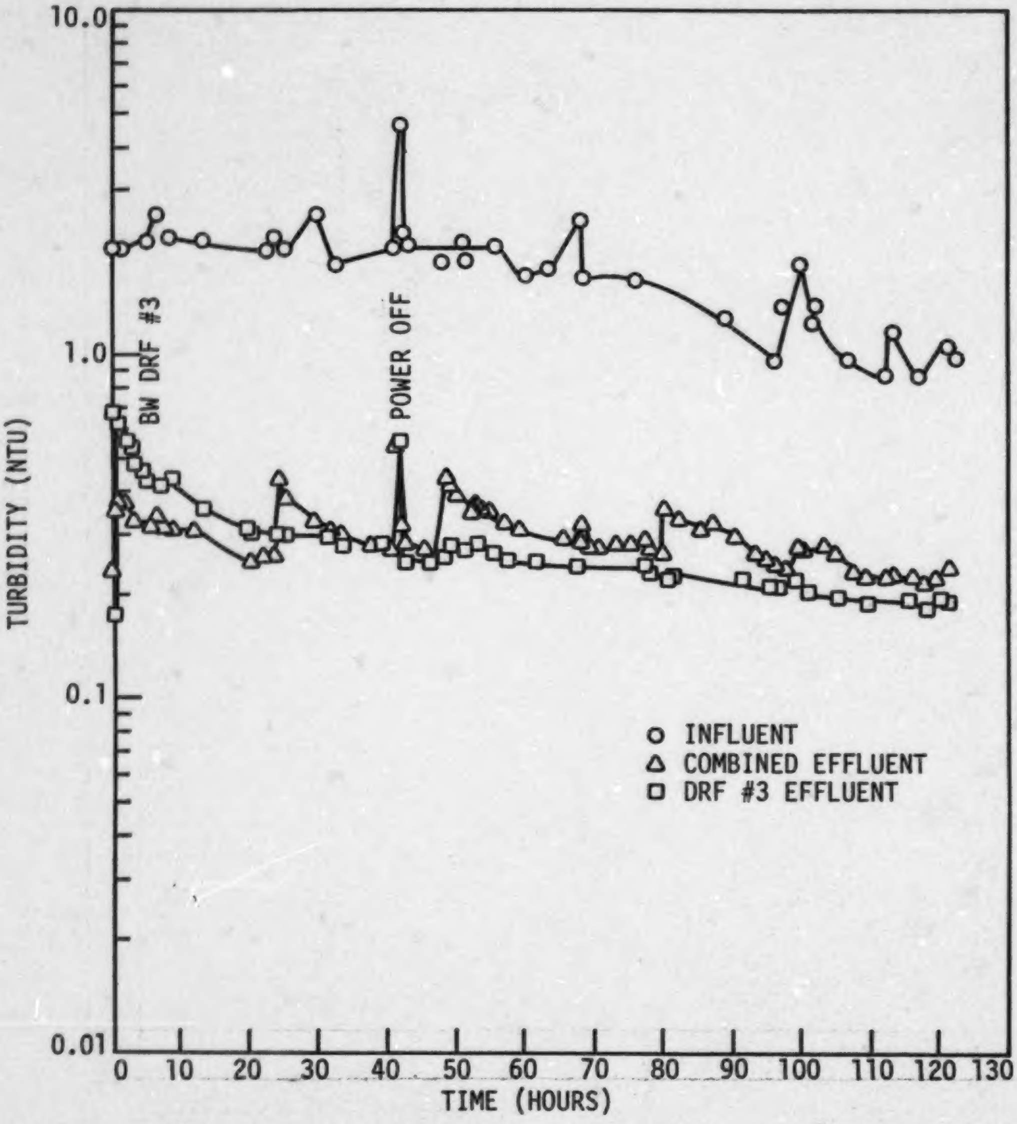


Fig. A-14. Turbidity for DRF during Run J-9 at 7.7 m/h (3.1 gpm/ft^2) using Cat-Floc T coagulant.

APPENDIX B: FILTRATION HEAD LOSS VS. TIME, PHASE II

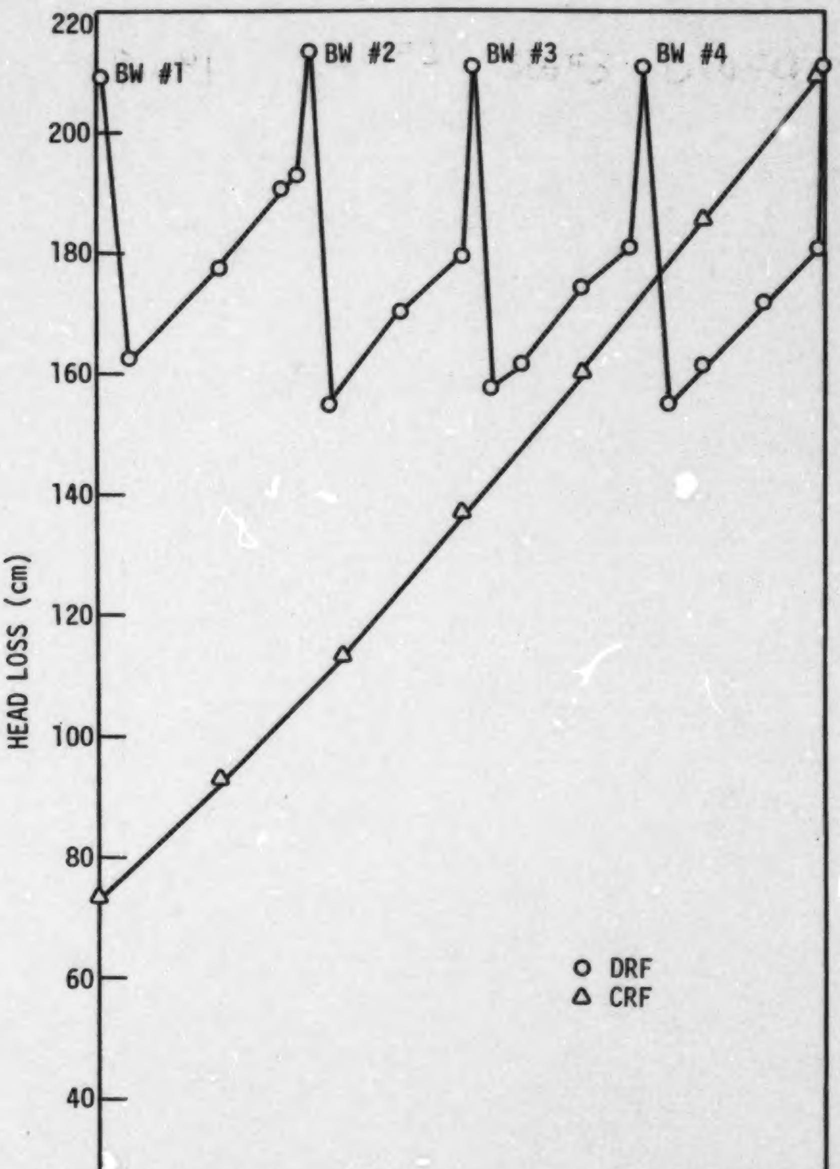


Fig. B-1. Head loss for CRF and DRF during Run J-1 at 13.2 m/h ($5.30\ gpm/ft^2$) using alum coagulant (feet = cm × 0.0328).

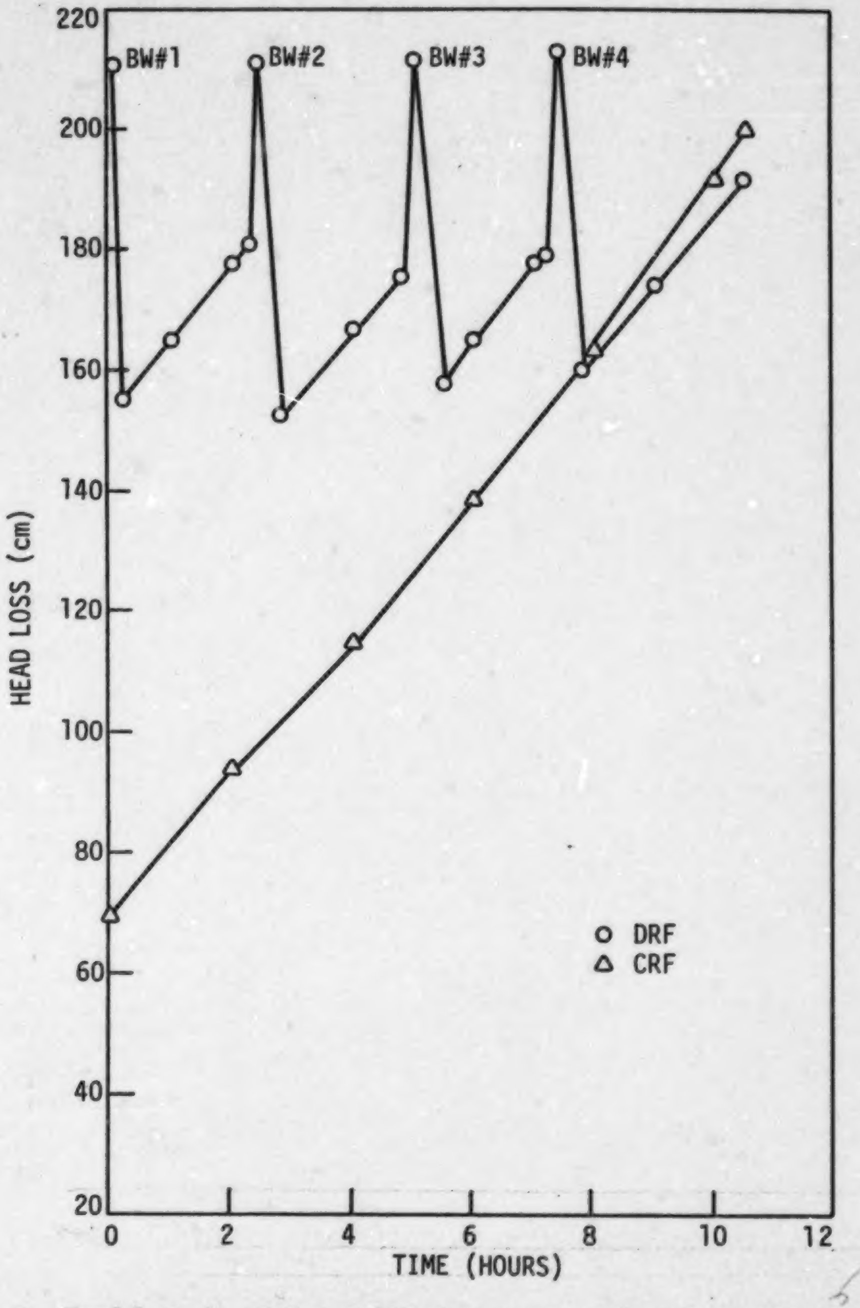


Fig. B-2. Head loss for CRF and DRF during Run J-2 at 13.5 m/h (5.42 gpm/ft^2) using alum coagulant (feet = cm × 0.0328).

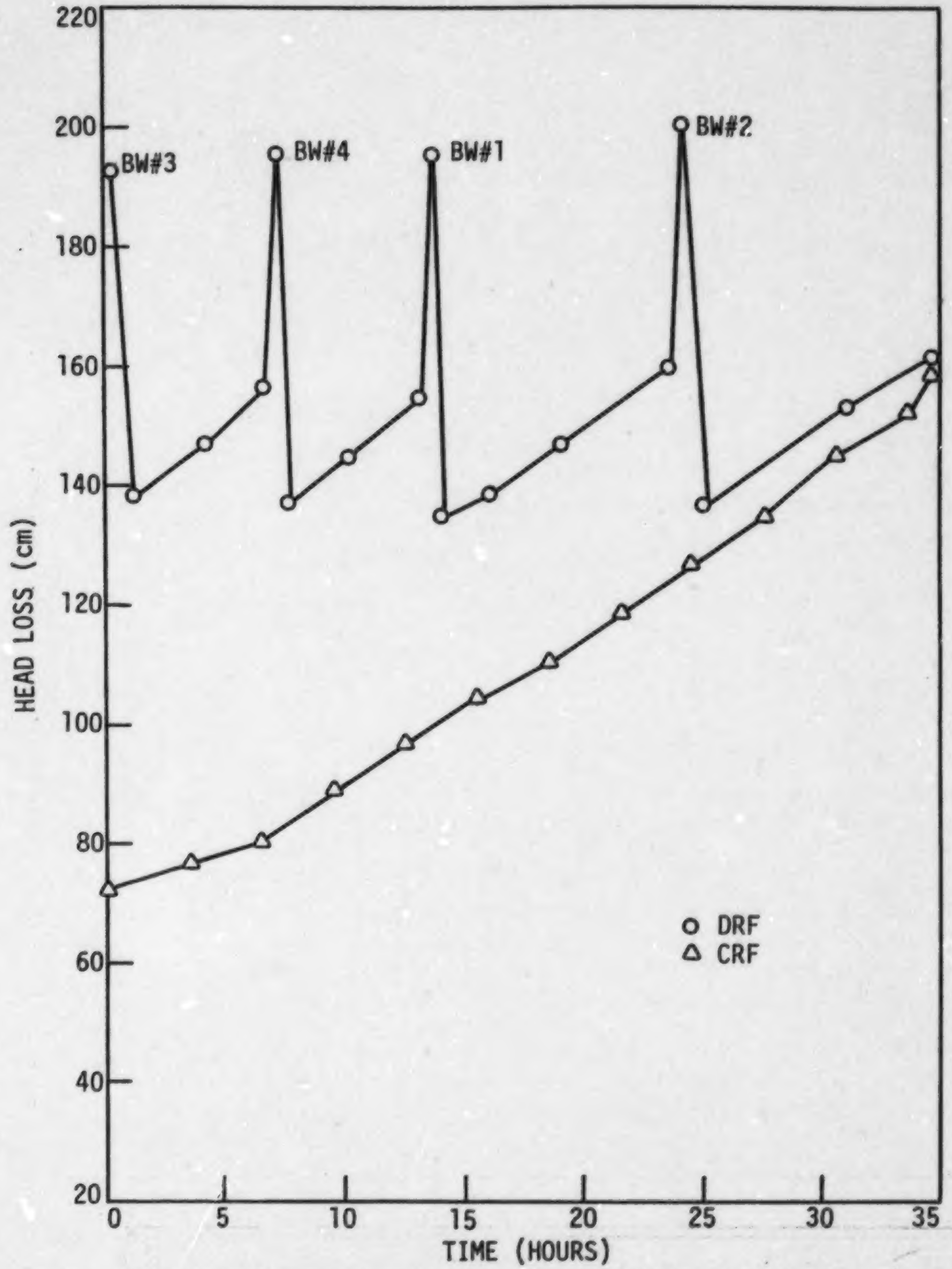


Fig. B-3. Head loss for CRF and DRF during Run J-4 at 13.4 m/h (5.38 gpm/ft^2) using Cat-Floc T coagulant (feet = cm × 0.0328).

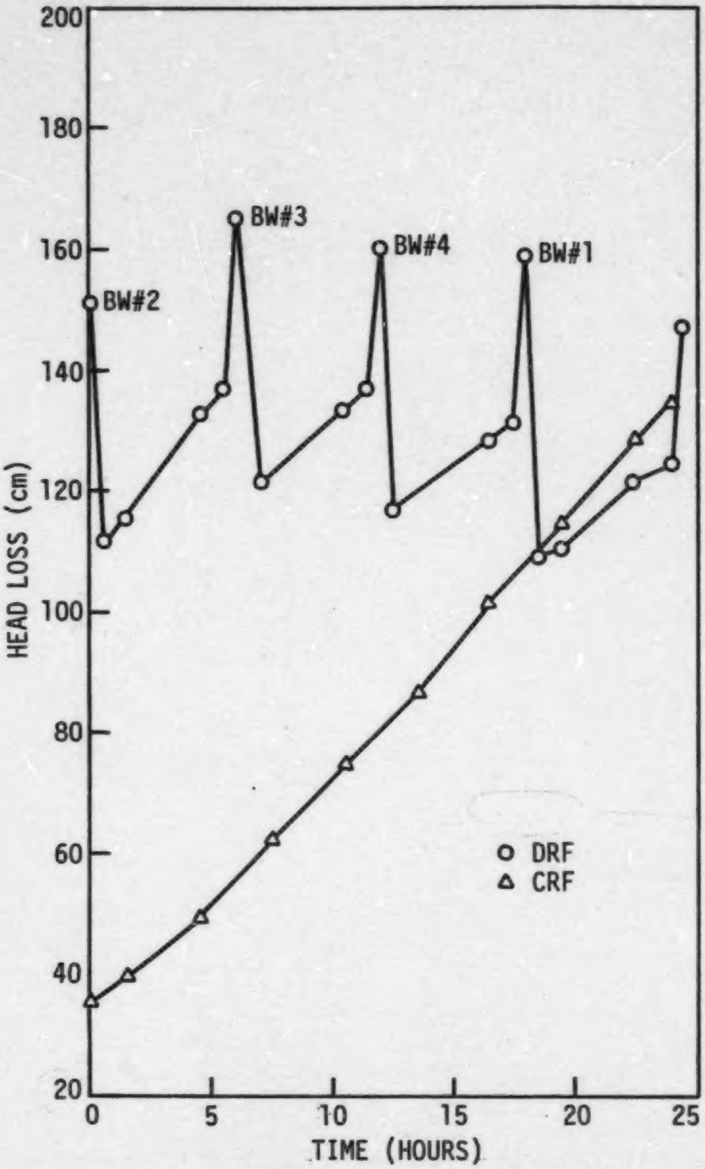


Fig. B-4. Head loss for CRF and DRF during Run J-6 at 7.7 m/h (3.1 gpm/ft^2) using alum coagulant (feet = cm × 0.0328).

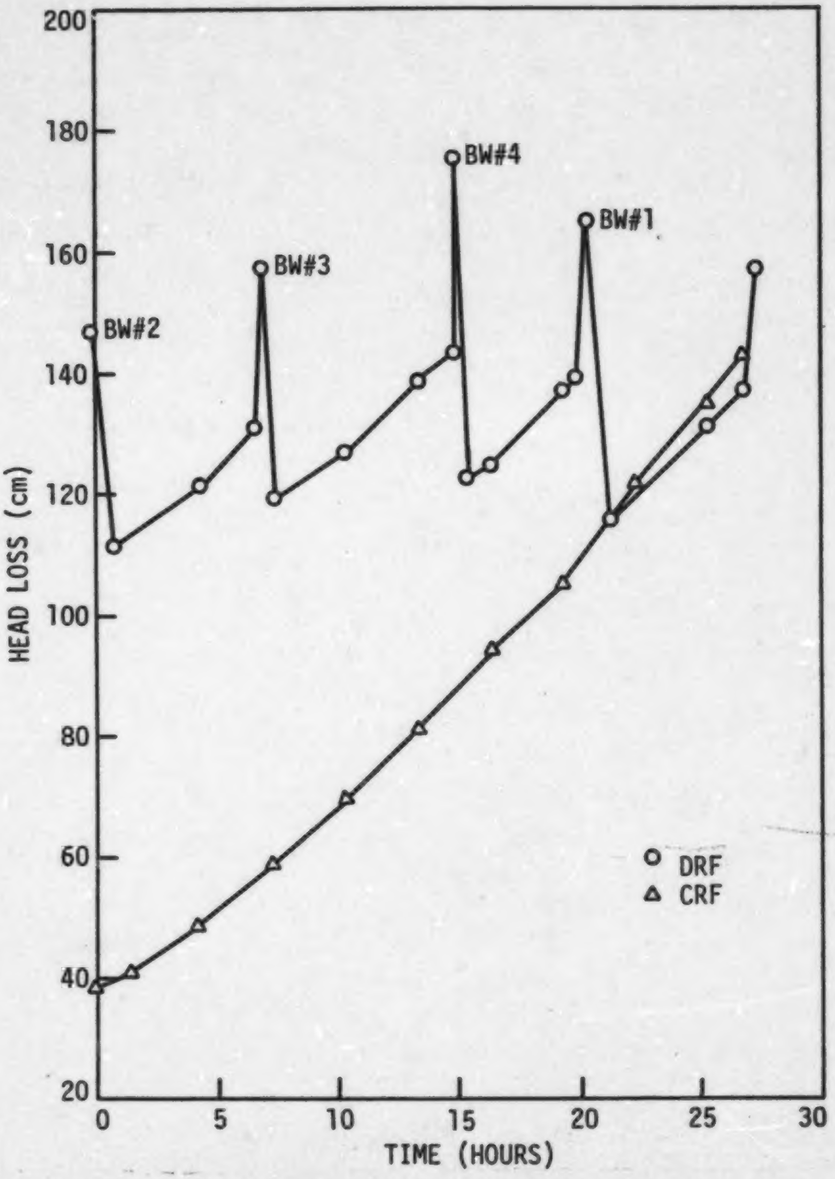


Fig. B-5. Head loss for CRF and DRF during Run J-7 at 7.6 m/h (3.1 gpm/ft^2) using alum coagulant (feet = cm × 0.0328).

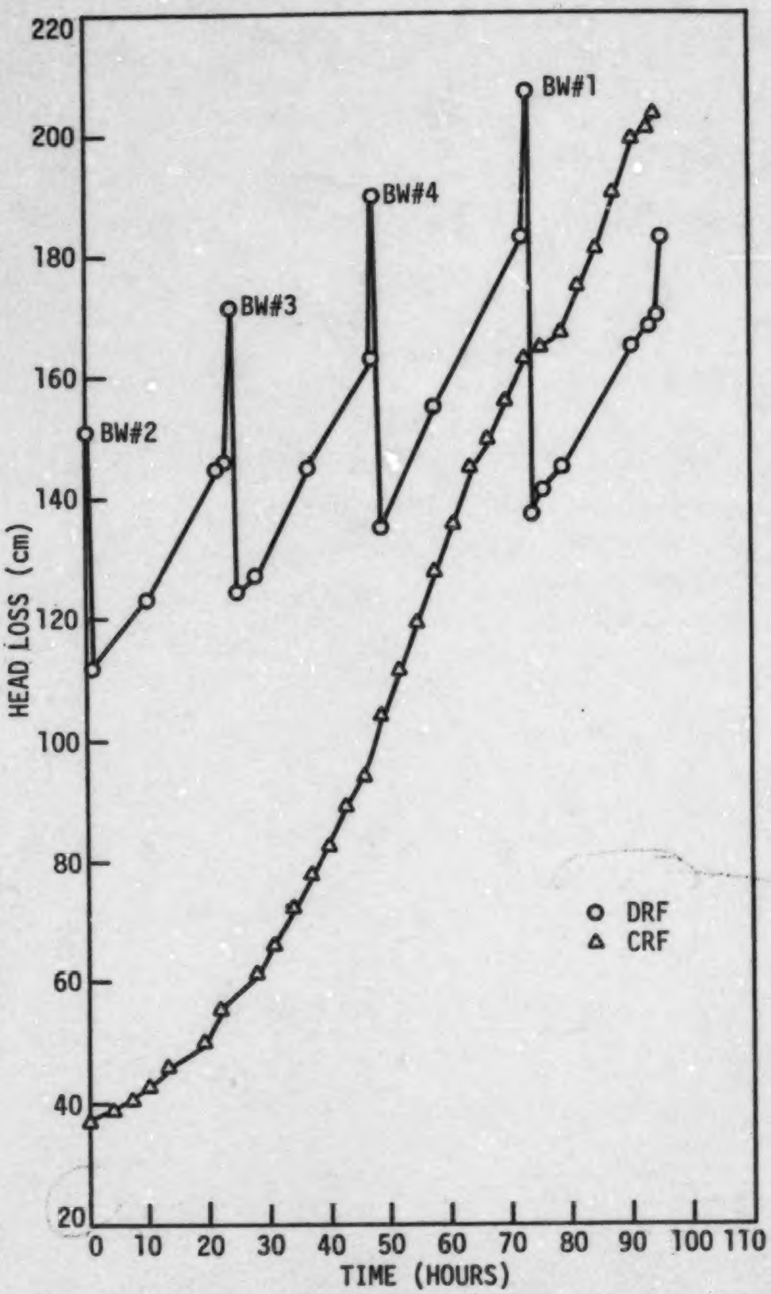


Fig. B-6. Head loss for CRF and DRF during Run J-8 at 7.7 m/h (3.1 gpm/ft^2) using Cat-Floc T coagulant (feet = cm × 0.0328).

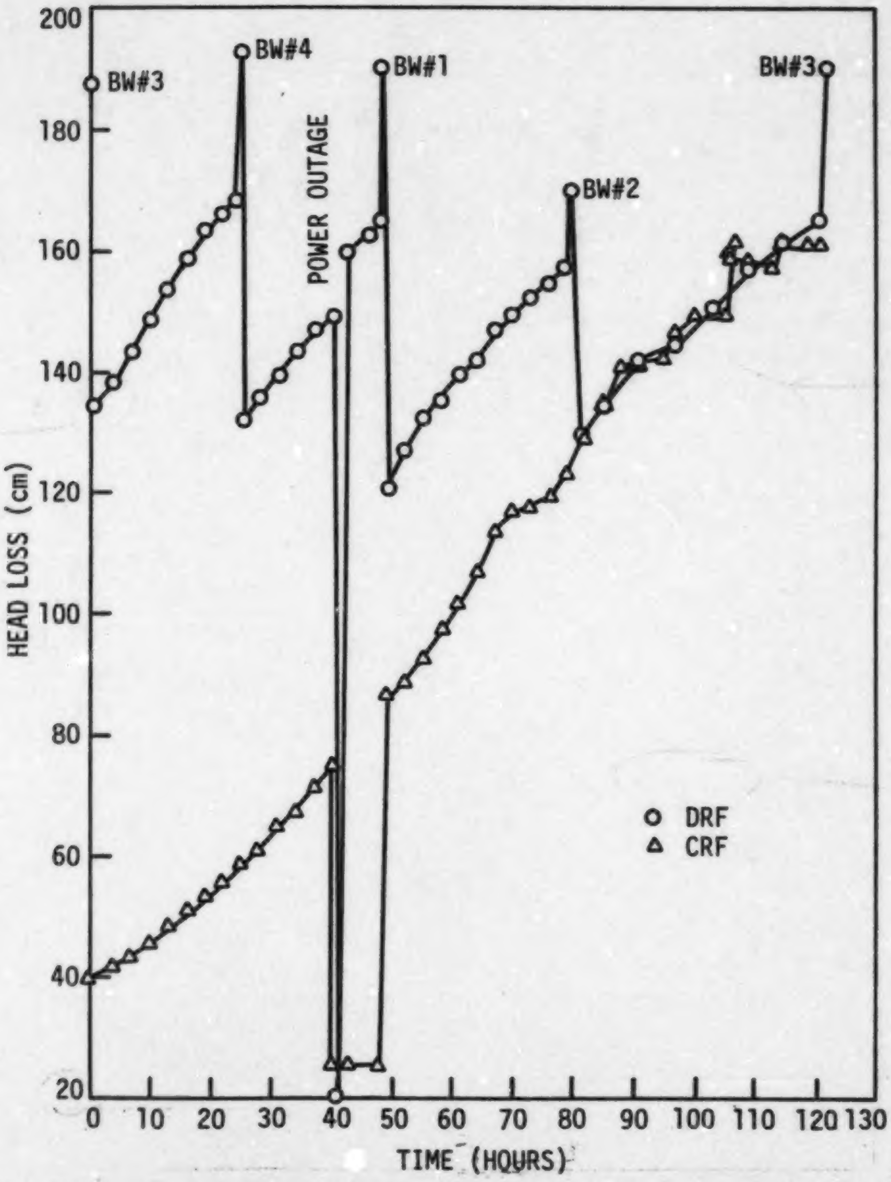


Fig. B-7. Head loss for CRF and DRF during Run J-9 at 7.7 m/h ($3.1\ gpm/ft^2$) using Cat-Floc T coagulant (feet = cm × 0.0328).

APPENDIX C: 7-12 MICRON PARTICLE COUNT AND TOTAL COLIFORM COUNT FOR RUN J-9, PHASE II

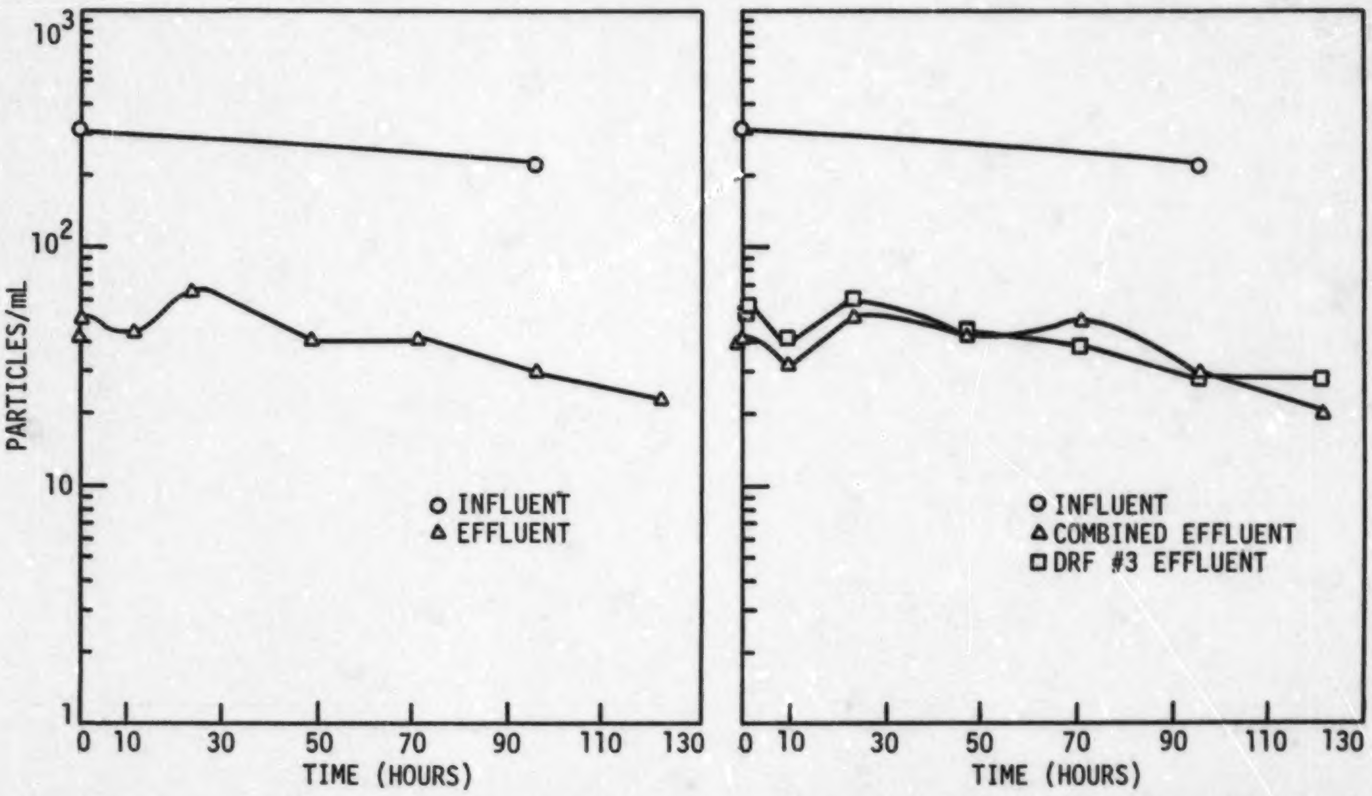


Fig. C-1. Seven to 12 µm particle count for CRF (left graph) and DRF (right graph) during Run J-9 at 7.7 m/h (3.1 gpm/ft^2) using Cat-Floc T coagulant.

C-1

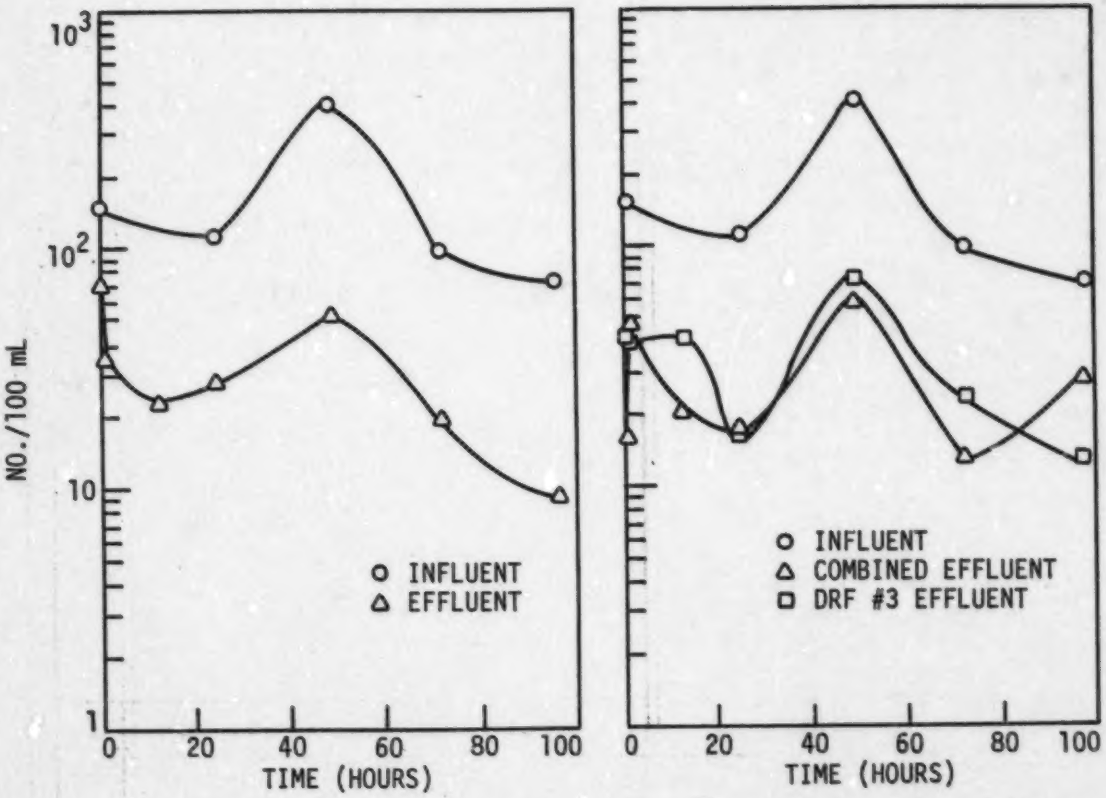


Fig. C-2. Total coliform count for CRF (left graph) and DRF (right graph) during Run J-9 at 7.7 m/h (3.1 gpm/ft^2) using Cat-Floc T coagulant.

APPENDIX D: INDIVIDUAL FILTRATION RATES VS. TIME DURING BACKWASH, PHASE II

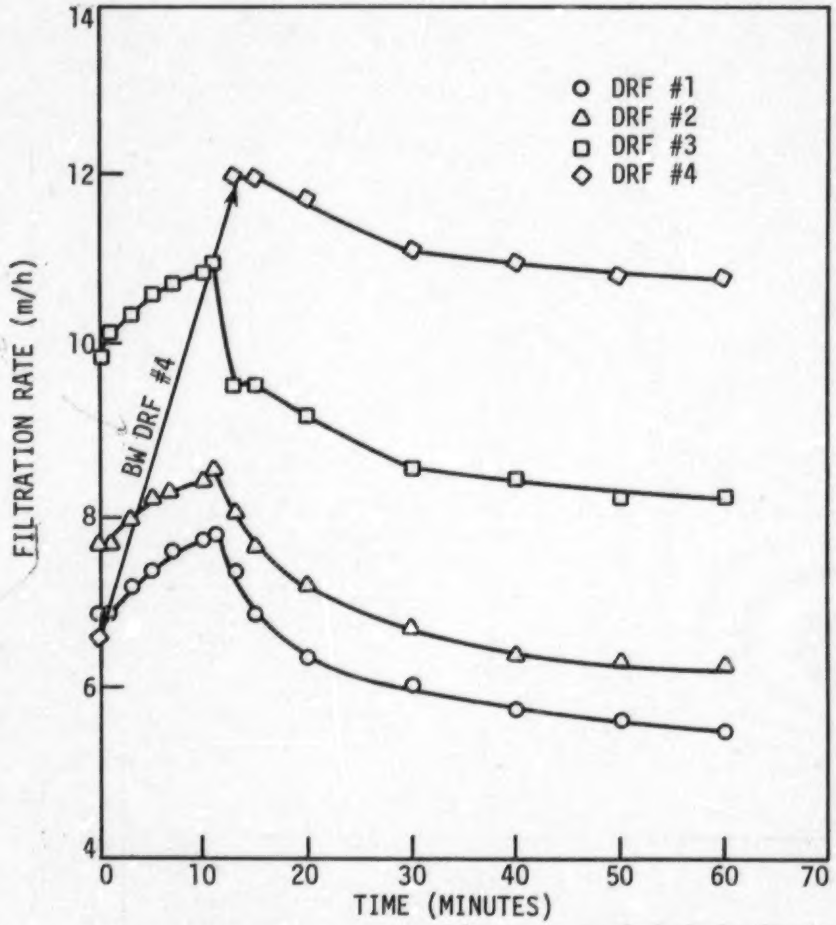


Fig. D-1. Changes in filtration rates for the four DRF during the backwash of DRF #4 on Sept. 24, 1982 (during Run J-8). Backwash begins at time = 0. Mean flow before backwash was 7.73 m/h (3.10 gpm/ft^2) and mean flow at time = 60 had returned to 7.68 m/h (3.08 gpm/ft^2) (gpm/ft^2 = m/h x 0.402).

D-1

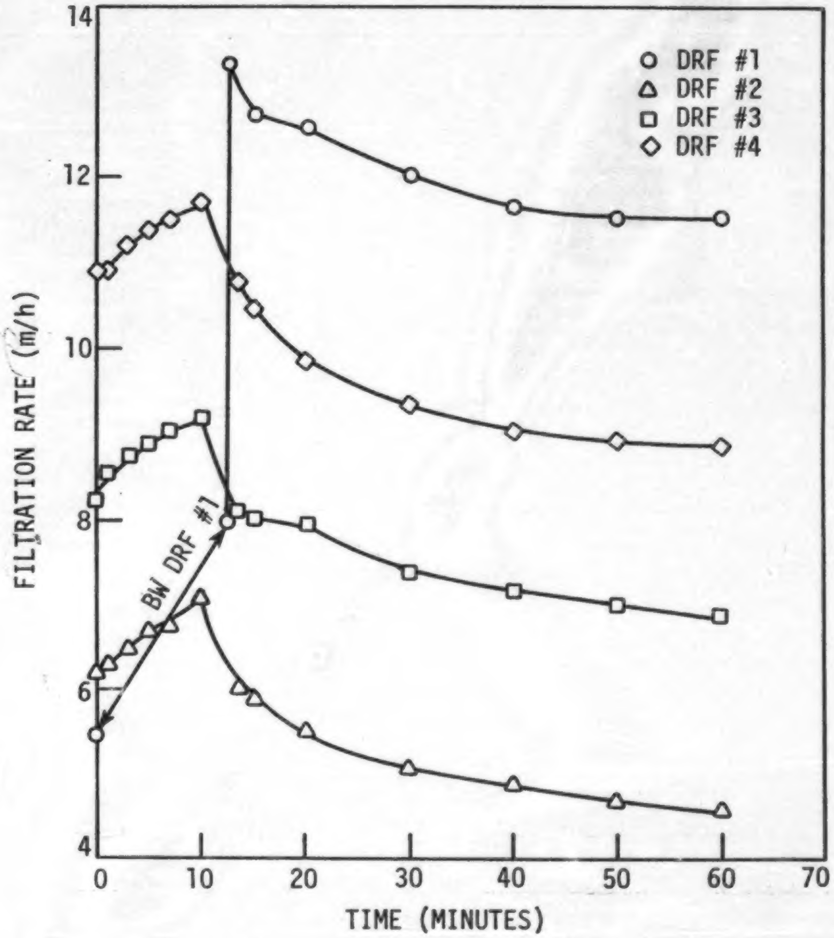


Fig. D-2. Changes in filtration rates for the four DRF during the backwash of DRF #1 on Sept. 25, 1983 (during Run J-8). Backwash begins at time = 0. Mean flow before backwash was 7.70 m/h (3.09 gpm/ft^2) and mean flow at time = 60 had returned to 7.95 m/h (3.19 gpm/ft^2) (gpm/ft^2 = m/h x 0.402).

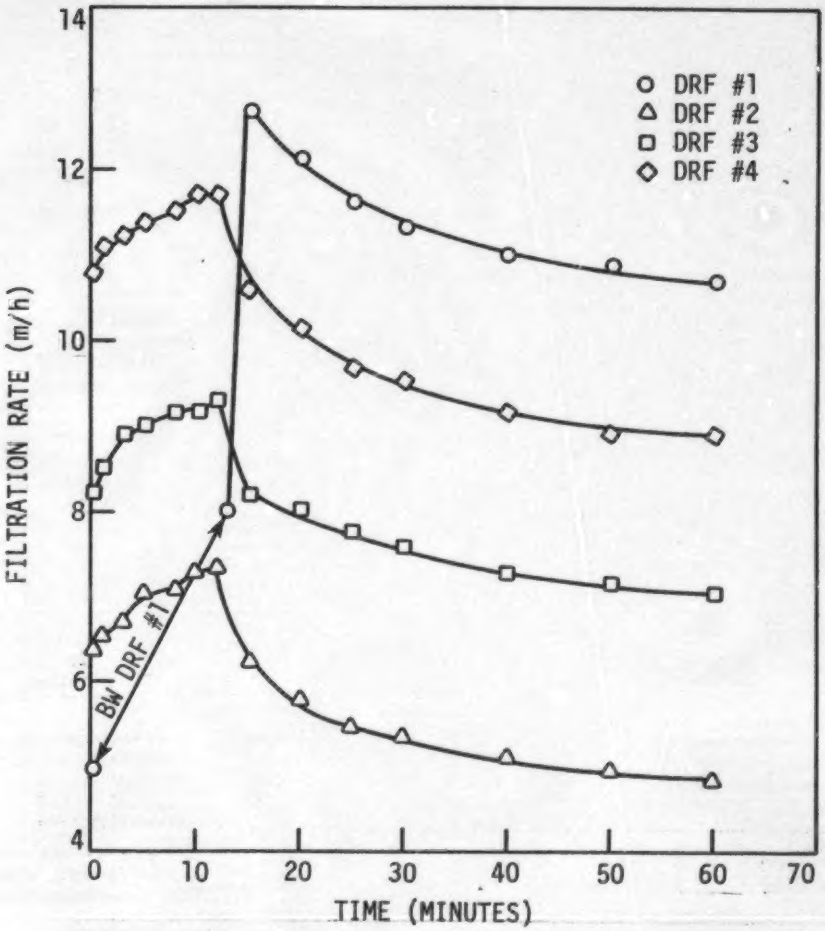


Fig. D-3. Changes in filtration rates for the four DRF during the backwash of DRF #1 on Sept. 29, 1983 (during Run J-9). Backwash begins at time = 0. Mean flow before backwash was 7.72 m/h (3.10 gpm/ft^2) and mean flow at time = 60 minutes had returned to 7.90 m/h (3.17 gpm/ft^2) (gpm/ft^2 = m/h x 0.402).